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PB-295 780

Treatment of Refinery Wastewater Using a Filtration-Activated Carbon System

BP Oil, Inc, Marcus Hook, PA

Prepared for

Robert S Kerr Environmental Research Lab, Ada, OK Source Management Branch

Mar 79

United States
Environmental Protection
Agency

Robert S. Kerr Environmental Research
Laboratory
Ada OK 74820

EPA-600/2-79-066
March 1979

Research and Development

PB 295780

EPA

Treatment of Refinery Wastewater Using a Filtration-Activated Carbon System

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TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/2-79-066	2.
3. RECIPIENT'S ACCESSION NO. PB295780	
4. TITLE AND SUBTITLE Treatment of Refinery Wastewater Using a Filtration-Activated Carbon System	5. REPORT DATE March 1979 (issuing date)
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Bruce A. McCrodden	8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS BP Oil Inc. Marcus Hook, PA 19061	10. PROGRAM ELEMENT NO. 1BB610C
	11. CONTRACT/GRANT NO. 12050 GXF
12. SPONSORING AGENCY NAME AND ADDRESS Robert S. Kerr Environmental Research Laboratory Office of Research and Development U.S. Environmental Protection Agency Ada, Oklahoma 74820	13. TYPE OF REPORT AND PERIOD COVERED Final 10/73 - 6/78
	14. SPONSORING AGENCY CODE EPA/600/15
15. SUPPLEMENTARY NOTES	

16. ABSTRACT

The objective of this report was to demonstrate the applicability of a dual media filtration-activated carbon adsorption system for the treatment of petroleum refinery wastewater. Constructed on a one-quarter acre plot, the capital cost of the wastewater treatment plant was $1,812,000 with an annual operating cost of $223,980.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Petroleum Refining Waste Treatment Industrial Wastes Activated Carbon Regeneration (engineering)	Wastewater Management	68 D 91 A

18. DISTRIBUTION STATEMENT Release to Public	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 91
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE PC-A05/MF-A01

EPA Form 2220-1 (9-73)

☆ U. S. GOVERNMENT PRINTING OFFICE: 1979-657-060/1629 Region No. 5-11

EPA-600/2-79-066
March 1979

TREATMENT OF REFINERY WASTEWATER USING A FILTRATION-ACTIVATED CARBON SYSTEM

by

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FOREWORD

The Environmental Protection Agency was established to coordinate administration of the major Federal programs designed to protect the quality of our environment.

An important part of the agency's effort involves the search for information about environmental problems, management techniques and new technologies through which optimum use of the nation's land and water resources can be assured and the threat pollution poses to the welfare of the American people can be minimized.

EPA's Office of Research and Development conducts this search through a nationwide network of research facilities.

As one of these facilities, the Robert S. Kerr Environmental Research Laboratory is responsible for the management of programs to; (a) investigate the nature, transport, fate and management of pollutants in groundwater; (b) develop and demonstrate methods for treating wastewaters with soil and other natural systems; (c) develop and demonstrate pollution control technologies for irrigation return flows; (d) develop and demonstrate pollution control technologies for animal production wastes; (e) develop and demonstrate technologies to prevent, control or abate pollution from the petroleum refining and petrochemical industries, and (f) develop and demonstrate technologies to manage pollution resulting from combinations of industrial wastewaters and industrial/municipal wastewaters.

This report contributes to the knowledge essential if the EPA is to meet the requirements of environmental laws that it establish and enforce pollution control standards which are reasonable, cost effective and provide adequate protection for the American public.

William C. Galegar

W.C. Galegar
Director
Robert S. Kerr Environmental
Research Laboratory

ABSTRACT

The objective of this project was to demonstrate the application for a dual media filtration-activated carbon adsorption system for total treatment of refinery wastewaters.

BP Oil, Inc.'s Marcus Hook Refinery has operated a wastewater treatment system consisting of dual media filtration for removal of oil and suspended solids followed by granular activated carbon adsorption for removal of dissolved organic material. Associated equipment includes backwash holding tanks, sludge thickners, two-stage centrifugation for oil-water-solids separation and a multiple hearth furnace for carbon regeneration.

The 2.2 MGD wastewater treatment plant has demonstrated average removals by the dual media filters of 58, 67, and 22 percent reduction for oil, suspended solids, and first stage ultimate oxygen demand, respectively. Average removals by the activated carbon absorbers have been 70, 27, 32, and 39 percent reduction for oil, first-stage ultimate oxygen demand, chemical oxygen demand, and phenol, respectively.

Constructed on a one-quarter acre plot, the capital cost of the wastewater treatment plant was $ 1,812,000 with an annual operating cost of $223,980.

This report was submitted in fulfillment of demonstration grant number 1205GXF by BP Oil, Inc. under the partial sponsorship of the U.S. Environmental Protection Agency. This report covers a period from October 1, 1973, to December, 1975, and work was completed as of June 1, 1978.

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ACKNOWLEDGMENT

The cooperation of the Standard Oil Company's (Ohio) Engineering and Research and Development groups is gratefully acknowledged for their participation in the design of the project and sampling program.

The aid of Mr. Leon H. Myers, U.S. Environmental Protection Agency, Robert S. Kerr Environmental Research Laboratory was particularly valuable during the project and reviewing the demonstration grant report.

The support of the Calgon Corporation Water Management Division is acknowledged for their evaluation of the activated carbon adsorption system and recommended systems improvements (Calgon report to BP Oil Corporation, Report No. C-850, dated June 13, 1974).

SECTION 1

INTRODUCTION

A 2.2 MGD* filtration-carbon adsorption Wastewater Treatment Plant was placed in operation at the Marcus Hook Refinery of BP Oil Inc., a subsidiary of the Standard Oil Company (Ohio), in March, 1973.

The Marcus Hook Refinery is a 150,000 BPD Class B Refinery located in Southeastern Pennsylvania. During the project period the refinery was modernized to take full advantage of its design capacity. Prior to December 1974, the refinery was operated at 105,000 BPD.

In 1969, a compliance schedule of 48 months was established to meet discharge standards prescribed by the Delaware River Basin Commission (DRBC). The initial effort toward achieving compliance was the evaluation of the existing API oil-water-solids Separator, through which all process wastewater flow is directed. Monitoring of API Separator influent and effluent first stage ultimate oxygen demand determined an average 68 percent removal, far below the DRBC's required 89.25 percent for process wastewater streams.

Accordingly, a project to determine the treatability of the API Separator effluent and a project to reduce the API Separator's hydraulic loading were undertaken. The latter project had as its basis an in-plant water use survey which concluded that a reduction in process wastewater flow to the API Separator could be accomplished by installation of a brine cooler; replacement of barometric condensers with surface condensers; segregation of sanitary wastes from the process wastewater stream; and further segregation of oily water and once through cooling water streams. The results of this project are evidenced by a reduction in the hydraulic loading from 3750 to 1700 GPM.

Treatability of the API Separator effluent was investigated through the operation of a bench scale activated sludge unit and an extended aeration pilot plant. With accumulated data as the basis, a preliminary biological treatment system design was prepared. The proposed full scale design required intermediate facilities for oil removal, two 369,000 gallon aeration basins, final clarifiers, an anaerobic digester, and both biological and oily sludge dewatering facilities.

Review of an estimated biological treatment capital cost of $2,500,000, an estimated annual operating cost of $220,500, biological treatment variability, land requirements, and excess sludge generation led to investigation of a filtration-activated carbon adsorption treatment system.

With filtration/adsorption pilot plant data as the basis, a preliminary filtration-carbon adsorption design was prepared. Comparison of an estimated capital cost of $2,000,000; an estimated annual operation cost of $179,000; and the reduced land area requirements, with the biological treatment preliminary design, led to the decision to construct a filtration adsorption wastewater treatment system.

This report covers the first two and one half years of operation of this system and is submitted in fulfillment of Demonstration Grant No. 12050GXF from the Research and Monitoring Division of the U.S. Environmental Protection Agency.

The objectives of this project were to:

1. Assess the feasibility of a filtration/adsorption treatment system for petroleum refinery process wastewater.
2. Evaluate performance of the system.
3. Determine capital and operating costs.
4. Determine economic feasibility of carbon regeneration.
5. Assess the treated effluent for reuse possibilities.

*Metric Conversion Table - p. 79

SECTION 2

CONCLUSIONS

1. The filtration/adsorption wastewater treatment system did not produce an effluent equal to design expectations.

2. The factors contributing to the poor performance of the treatment system were:

   A. The waste load on the filters/adsorbers increased over that experienced during the pilot plant operation.

   B. A 40% decrease in the adsorptive capacity of the regenerated carbon was observed following 18 months of operation.

   C. A change occurred in the wastewater characteristics as evidenced by a decrease in the theoretical carbon capacity.

3. The carbon adsorber design flow rate of 1500 GPM could not be maintained due to plugging of the effluent septums by carbon fines.

4. The static bed activated carbon pilot plant did not provide an adequate basis for design of a full scale pulse bed system.

5. Carbon losses stabilized to 6% per regeneration cycle.

6. The production of sulfide across the carbon columns was not only the result of bacterial action but was also a function of influent characteristics.

7. The dual media filters demonstrated consistent removal of suspended solids and oil during 18 months of operation. The filters' removal efficiency decreased as media was gradually lost from the filter vessels.

8. The solids dewatering system created a recycle of solids and oil to the API Separator influent which in turn impacted on all down stream units.

SECTION 3

RECOMMENDATIONS

1. All wastewater streams to be treated in the full scale units should be included in the pilot plant influent.

2. Activated carbon pilot plants should be dynamic systems that model the full scale unit. Pilot scale regeneration systems should be operated to obtain scale up parameters and assess losses in adsorptive capacity.

3. Further investigation as to the production of sulfide during treatment of refinery wastewater should be undertaken.

4. Further investigation as to the optimum regeneration furnace temperature profiles should be undertaken.

5. The design of activated carbon systems should provide adequate storage capacity for regenerated and spent carbon to allow for regeneration furnace shutdowns.

6. The design of activated carbon systems should include facilities for removing fines from the regenerated carbon.

7. Pulsed bed carbon adsorption columns should be designed as pressure vessels to permit increased flow rates.

SECTION 4

PROCESS DESCRIPTION

FILTRATION

Filtration is the process of "straining" suspended, insoluble matter from a liquid stream. Various mechanisms are responsible for the translation of suspended particles from the main stream of flow to the filter grains in a deep-bed system. There are also various forces which will retain the particle once it is brought into contact with the filter medium.

It appears that the random movement of suspended particles in a flow path is mainly responsible for transporting the particles either directly to the grains or close enough to the grain surfaces for other forces to become effective. Removal mechanisms can be discussed as a function of particle diameter for small particles, with diameters one micron and smaller, random movement dominates in bringing the particles to, or close to, the grain surfaces. Van der Waals forces will accomplish capture once particles are within 0.05 to 0.1 micron range, outside of which gravity forces will dominate. For particles of very small diameter, Brownian diffusional forces become increasingly important in the last stage of the contact. In most cases, retention is due to Van der Waals forces, but for fine particles with positive charges, electrokinetic forces will be responsible. Chemisorption, ion exchange, and chemical bonding will operate in the atomic and molecular ranges.

For large particles with diameters above 50 microns, direct contact predominates, especially for particles to be captured on multiple contact sites. Sedimentation is of secondary importance. When a particle is captured on a multiple-contact site, it reduces the passageway and causes capture of progressively finer particles. On the bottom side of the grains, Van der Waals forces may contribut to the capture mechanism. On the bottom side of the filter grains, Van der Waals forces are strongly opposed by the earth's gravitational pull. On the top side of the grains, these forces are additive and more sediment is expected to have much electrokinetic potential; electrokinetic forces, therefore, are not important in retention. Friction forces and fluid pressure may help retain particles captured in multiple-contact sites.

For medium size particles with diameters one to 50 microns most of the mechanisms and forces applicable to the other two classes will have relevance. The larger particles in this class may still be subject to capture by direct interception on multiple-contact sites in fine filter media (grain diameter = 0.3 millimeter or less). The majority, however, will be captured by a combination of random movement and sedimentation, with Van der Waals forces increasingly more important as the particle size and density decrease.

Friction and fluid pressure play a part in connection with the larger particles captured on multiple-contact sites. Electrokinetic adhesion forces may operate in the small particle range in the case of opposing charges, but, in the bulk of this class, Van der Waals adsorption and gravity forces predominate. On the upper surfaces these two forces are additive; on the bottom surfaces, they oppose each other. Since the superiority of Van der Waals forces over gravity decreases with the inverse second power of the particle diameter, larger particles will have a greater preference for the upper surfaces than the small particles. It is also to be expected that the particle adhesion forces will diminish with each layer of particle deposition, since these layers are usually not as dense as the filter medium. This fact again tends to increase the relative importance of gravity over Van der Waals forces. In other words, it results in a thicker sediment layer on the top than on the bottom surfaces with increasing particle size.

The substances removed during filtration are distributed irregularly over the grain surfaces and are not dislodged by the passing fluid under normal operating conditions. Interstices, however, are narrowed down by accumulating deposits and some may be completely closed. Particles entering pores still open are then transported deeper into the bed, until they reach grain sites still able to accept them. Only when particles fail to find such sites do they pass into the effluent.

ACTIVATED CARBON ADSORPTION

Activated carbon removes organic contaminants from water by the process of adsorption or the attraction and accumulation of one substance on the surface of another. In general, high surface area and pore structure are the prime considerations in adsorption of organics from water; whereas, the chemical nature of the carbon surface is of relatively minor significance. Granular activated carbons typically have surface areas of 500-1,400 square meters per gram. Activated carbon has a preference for organic compounds and, because of this selectivity, is particularly effective in removing organic compounds from aqueous solution.

Much of the surface area available for adsorption in granular carbon particles is found in the pores within the granular carbon particles created during the activation process. The major contribution to surface area is located in pores of molecular dimensions. A molecule will not readily penetrate a pore smaller than a certain critical diameter and will be excluded from pores smaller than the designated critical diameter.

Activated carbon is manufactured by a process consisting of raw material dehydration and carbonization followed by activation. The starting material is dehydrated and carbonized by slowly heating in the absence of air. The activated carbon used in this project was made from bituminous coal.

Adsorption by activated carbon involves the accumulation or concentration of substances at a surface or interface. Adsorption is a process in which matter is extracted from one phase and concentrated at the surface of another, and is therefore termed a surface phenomenon. Adsorption from wastewater onto activated carbon can occur as a result of two separate properties of the wastewater-activated carbon system, or some combination of the two: (1) the low solubility of a particular solute in the wastewater and (2) a high affinity of a particular solute in the wastewater for the activated carbon. According to the most generally accepted concepts of adsorption, this latter surface phenomenon may be predominantly one of electrical attraction of the solute to the carbon, of Van der Waals attraction, or of a chemical nature.

There are essentially three consecutive steps in the adsorption of dissolved materials in wastewater by granular activated carbon. The first step is the transport of the solute through a surface film to the exterior of the carbon. The second step is the diffusion of the solute within the pores of the activated carbon. The third and final step is adsorption of the solute on the interior surfaces bounding the pore and capillary spaces of the activated carbon.

There are several factors which can influence adsorption by activated carbon, including: (1) the nature of the carbon itself; (2) the nature of the material to be adsorbed, including its molecular size and polarity; (3) the nature of the solution, including its pH; and (4) the contacting system and its mode of operation.

THERMAL REGENERATION

Thermal regeneration of granular carbon consists of three steps: (1) drying; (2) pyrolysis of adsorbates; and (3) activating by oxidation of the carbon residues from decomposed adsorbates. Drying is accomplished at 212°F, baking between 212 and 1500°F, and activating at carbon temperatures above 1500°F. All

of these steps are carried out in a direct-fired hearth furnace. Time, temperature and atmosphere are the controllable variables for regeneration. Free oxygen must be controlled by the addition of steam in the lower hearths of the furnace to avoid burning of the granular carbon itself.

CENTRIFUGATION

Centrifugation may be defined as sedimentation under the influence of forces greater than gravity. A centrifuge can clarify, classify, or separate components of a given stream as a function of the difference in the component's specific gravities. The disc and scroll type centrifuges are discussed in this report. The fundamental difference in these types is the method by which components are collected in, and discharged from the bowl. The method of discharge determines the size and nature of the particles which are suitably collected and handled in each of the centrifuges.

Within a centrifuge, centrifugal force acts on a suspended particle, causing it to settle through the liquid component. By rotating the settling vessel at high speed, the settling forces acting on a particle can be increased by several orders of magnitude.

The degree of removal within the centrifuge is both a function of the average retention time and the effective centrifugal force acting on the component to be separated. Separation of the components is effected when the settling velocity imparted to them by the centrifugal force exceeds the overflow velocity of the suspending liquid. The rate of separation is dependent on the differential density between the individual components.

Liquid viscosity relationships apply to centrifugal separation. As the temperature of the liquid increases, the viscosity decreases, and thus the rate of subsidence of the particle to be separated increases.

Particle distribution or concentration is interrelated with particle size and shape, and more generally affects the concentration of the scroll centrifuge cake. The nature and compressibility of the solids also affect the concentration of the cake.

SECTION 5

PILOT PLANT STUDY

DUAL MEDIA FILTRATION

A pilot plant filtration/adsorption treatment of API Separator effluent was investigated over a period of six weeks in August and September, 1970. The equipment used for this pilot plant study was a dual media filter with a cross-sectional area of one square foot. The unit was square, 17 feet in overall height and had one face of transparent plexiglass. The filtering media consisted of 5 ft. of sand and 2.5 ft. of anthrafilt supported by 16 inches of gravel. A schematic of the filtration pilot plant is shown in Figure 1.

API Separator effluent was fed to the filtration pilot plant. The filter effluent was discharged to a 55-gallon drum for use as feed to the activated carbon columns. The filter was backwashed with influent water when the pressure drop reached 13.5 psi. The backwash procedure consisted of an initial air scour followed by a water rinse.

The influent to and effluent from the sand filter were composited during each run and were analyzed for oil, phenol, suspended solids, BOD_5, and TOC. A total of 37 runs were conducted using flow rates of 12 to 18 gal./min./sq. ft. with most runs at the lower end of the range.

The first five runs were conducted using only 5.5 ft. of sand. The results indicated that performance was good but that the filter runs were quite short, approximately 5 hours. Therefore, to improve filter runs, 2.5 feet of anthrafilt were added. Filter runs were conducted at 12 gpm./sq. ft. filtration rates, 16 gpm/sq. ft. and 18 gpm/sq. ft. In the sand anthrafilt runs, performance of the filter did not change significantly when the filtration rate was varied from 12 to 18 gpm/sq. ft.

Use of coagulants at dosages of 1 to 5 mg/L did not improve the performance of the sand filter.

The performance data for the filtration of API Separator effluent is presented in Table 1. Average removals of suspended solids, BOD_5, TOC, and oil were 77, 42, 48, and 79 percent

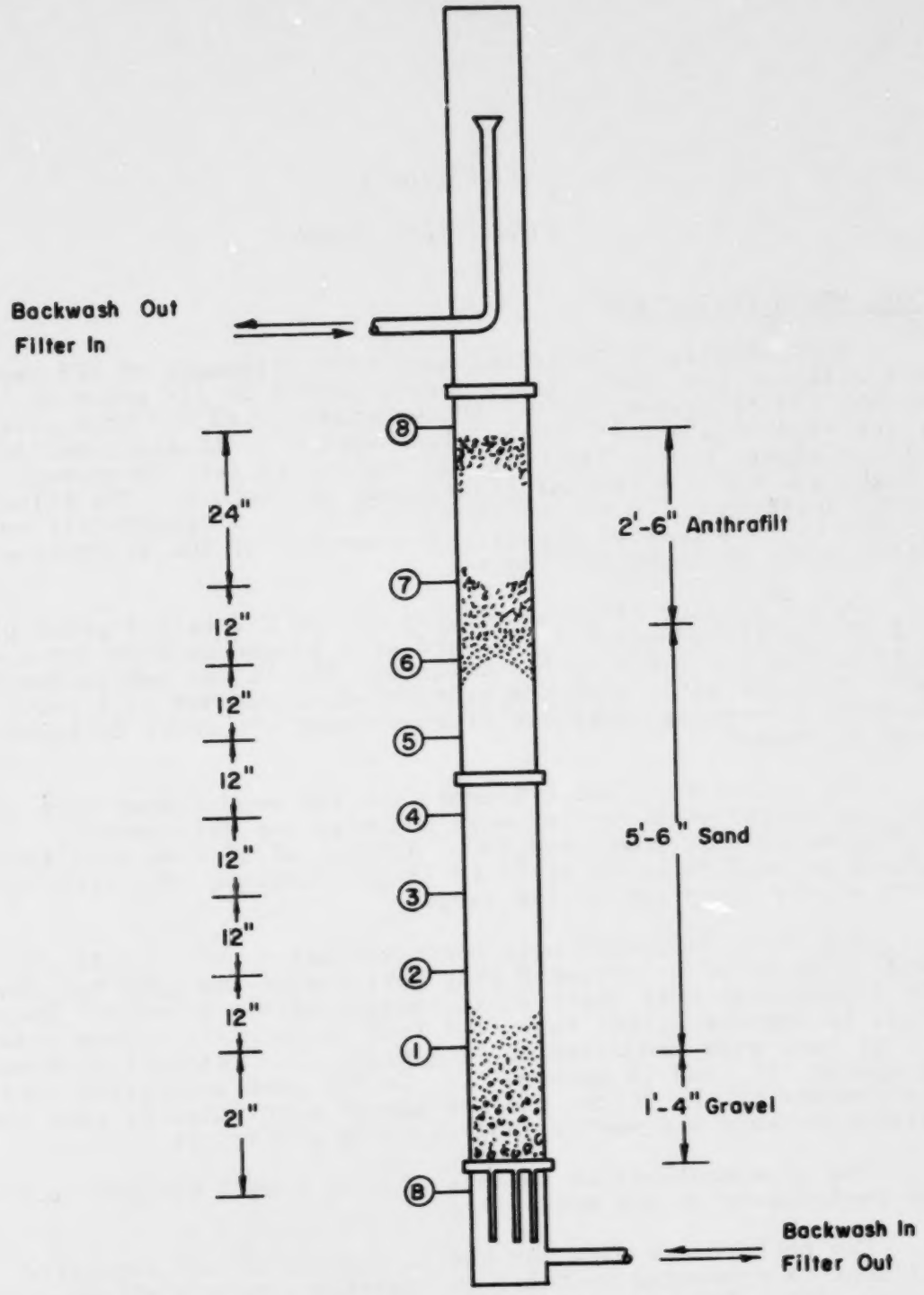


Figure 1. UHR test filter Functional diagram

TABLE 1

Filtration Pilot Plant Data

Date	Run No.	Suspended Solids mg/l Inf.	Suspended Solids mg/l Eff.	Suspended Solids Percent Removal	BOD_5 mg/l Inf.	BOD_5 mg/l Eff.	BOD_5 Percent Removal	TOC mg/l Inf.	TOC mg/l Eff.	TOC Percent Removal	Oil mg/l Inf.	Oil mg/l Eff.	Oil Percent Removal	Phenol mg/l Inf.	Phenol mg/l Eff.	Phenol Percent Removal
8/13/70	7	20	21	--	54	35	35	116	45	61	--	--	--	--	--	--
8/13/70	8	70	5	93	--	--	--	137	45	67	178	17	90	0.19	0.35	--
8/14/70	9	90	14	84	87	24	72	68	38	44	--	--	--	--	--	--
8/17/70	14	35	7	80	73	29	60	137	50	64	36	7.5	79	0.80	0.76	5
8/18/70	16	26	3	89	84	54	36	81	61	25	48	13.6	72	11.6	9.8	10
8/19/70	18	20	4	80	30	20	33	79	48	40	35	9.5	73	--	--	--
8/20/70	20	10	5	50	66	39	41	77	41	47	51	11	80	0.83	0.79	4
8/21/70	22	40	5	88	45	33	27	39	33	2	58	10	83	4.7	3.7	21
8/23/70	26	55	20	64	--	--	--	77	39	49	--	--	--	--	--	--
8/24/70	28	40	15	63	83	49	41	46	41	11	44	11	75	7.5	6.8	10
8/25/70	29	25	5	80	68	29	57	56	26	53	56	6.9	88	2.0	2.3	--
8/26/70	30	15	10	33	45	34	25	45	34	20	50	17.2	66	4.4	4.8	--
8/29/70	31	25	5	80	50	35	30	35	25	29	44	11	75	4.1	3.0	27
8/31/70	36	26	6	77	48	30	38	50	36	28	51	8.2	85	3.9	4.2	--
9/01/70	37	25	5	80	53	35	34	70	35	50	50	11	78	3.6	3.5	--
9/02/70	38	28	12	57	60	53	12	--	--	--	60	14.2	77	5.7	5.3	7
9/03/70	39	20	6	70	55	27	50	--	--	--	71	11.2	85	0.8	0.7	12
Average		35	8	77	60	35	42	74	40	48	60	14.2	79	3.85	3.54	8

respectively.

The filter runs varied from 12 to more than 24 hours, depending on the incoming TOC concentration, i.e., the higher the TOC concentration, the shorter the filter run.

The release of contaminants from the filter occurred during backwashing at 30 gal./min./sq. ft. An analysis of the oil and suspended solids content of the pilot plant filter backwash is presented in Table 2. The results indicate that a 6-minute backwashing was sufficient to clean the filter.

The average pilot plant influent and effluent values are summarized below:

Filtration Pilot Plant Average Performance

Oil			Suspended Solids			BOD_5		
mg/l		Percent	mg/l		Percent	mg/l		Percent
Inf	Eff	Removal	Inf	Eff	Removal	Inf	Eff	Removal
60	11	79	35	8	77	60	35	42

ACTIVATED CARBON ADSORPTION

The carbon columns used in the study were 5 inch diameter plexiglass columns. A schematic of the adsorption pilot plant is shown in Figure 2. Four columns were used in series; the first column was filled with 3 feet of carbon; the remaining columns contained 5 feet, giving a total carbon depth of 18 feet. The carbon used during the study was 8 x 30 mesh Filtrasorb 300. The sand filter effluent was fed to the carbon column at a rate of 0.5 gal./min. giving a rate of 3.6 gal./min./sq. ft. and a contact time of 36 minutes. A total of approximately 11,000 gallons of wastewater was passed through the carbon. The effluent from the carbon columns was composited to correspond to one complete sand filter run and analyzed for oil, phenol, suspended solids, BOD_5 and TOC. In addition, grab samples of effluent from carbon columns one and four were periodically analyzed for TOC.

The adsorption isotherm for filtered API Separator effluent at ambient temperature is shown in Figure 3. The results indicate that an effluent TOC of 3 mg/l could be obtained by carbon adsorption. The intersection of isotherm at the initial TOC concentration of 36 mg/l gives the theoretical capacity of the carbon when it is in equilibrium with the influent concentration. For this particular wastewater, the theoretical capacity was 0.3 lbs. of TOC adsorbed per pound of carbon. This is equivalent to 1 pound of carbon exhausted per 1,000 gallons of wastewater treated.

Approximately 11,000 gallons of wastewater were treated

TABLE 2

Summary of Analyses of Sand Filter Backwash Water[1]

Run No.	Time After Initiating Backwash, Mins.	mg/l Oil	mg/l Suspended Solids
17	0.5	14,500	7,550
	2.0	2,175	640
	3.0	830	596
	4.0	430	524
	6.0	175	102
28	0.5	23,000	12,050
	1.5	8,500	4,550
	2.5	1,155	1,140
	4.0	412	220
	7.0	72	40

[1]Backwash procedure consists of air and water scrub for 5 minutes followed by water rinse at 30/gal/min/ft.[2]

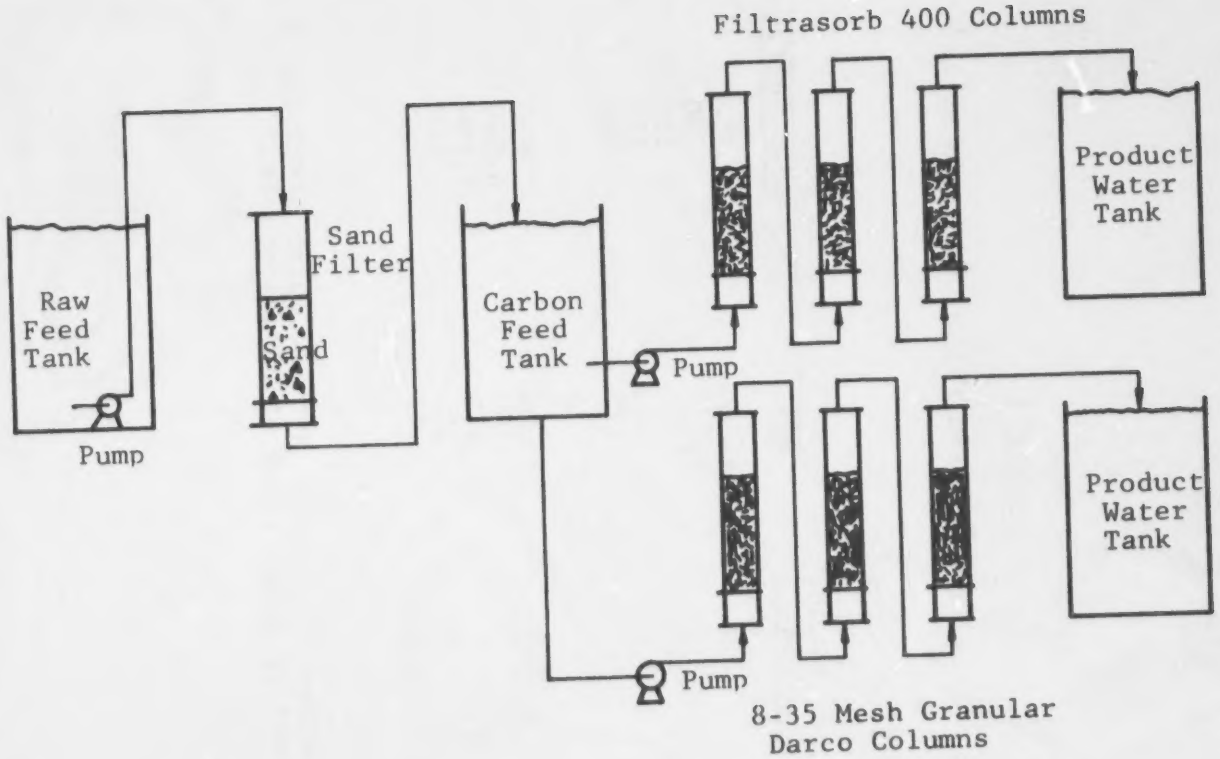


Figure 2. Laboratory adsorption system

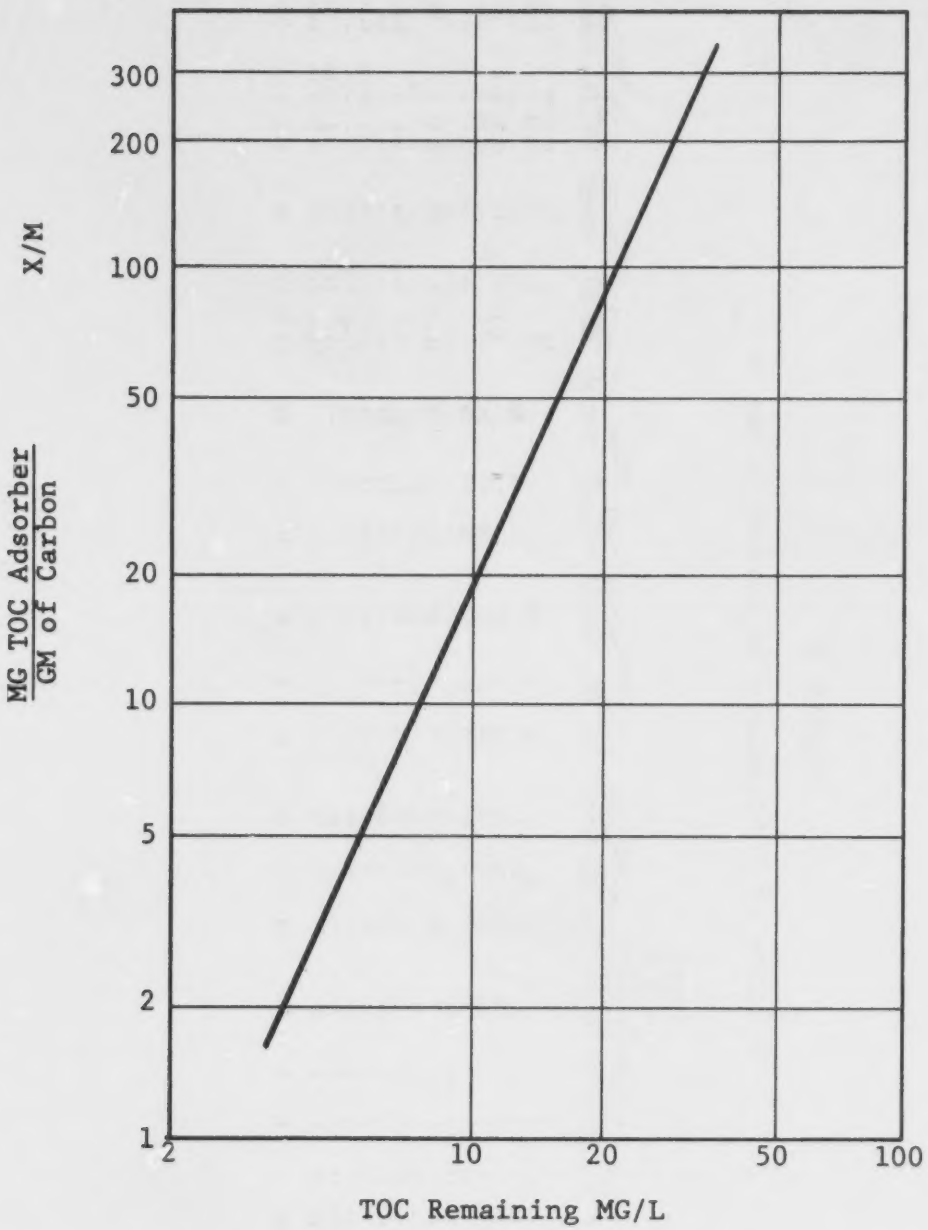


Figure 3. Carbon adsorption isotherm of filtered waste water

TABLE 3

Summary of Performance of Carbon Column 4

Date	Run No.	Suspended Solids mg/l Inf.	Suspended Solids mg/l Eff.	Suspended Solids Percent Removal	BOD_5 mg/l Inf.	BOD_5 mg/l Eff.	BOD_5 Percent Removal	FSUOD mg/l Inf.	FSUOD mg/l Eff.	FSUOD Percent Removal	TOC mg/l Inf.	TOC mg/l Eff.	TOC Percent Removal	Oil mg/l Inf.	Oil mg/l Eff.	Oil Percent Removal	Phenol mg/l Inf.	Phenol mg/l Eff.	Phenol Percent Removal
8/13/70	8	5	1	80	--	--	--	--	--	--	--	--	--	17	2.3	87	0.35	0	100
8/17/70	14	7	3	57	29	4	86	50	5	90	--	--	--	7.5	1.3	83	0.76	0	100
8/18/70	16	3	1	67	54	6	89	--	--	--	61	12	80	--	--	--	--	--	--
8/19/70	18	4	2	50	20	4	80	46	6	87	--	--	--	9.5	0.9	91	0.86	0.1	89
8/20/70	20	5	1	80	39	4	90	84	15	81	41	17	59	--	--	--	0.79	0	100
8/21/70	22	5	5	0	33	6	81	46	17	63	33	14	58	--	--	--	--	--	--
8/23/70	26	20	10	50	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
8/24/70	28	15	15	0	49	5	90	71	8	89	41	13	68	11	2	82	6.8	0	100
8/25/70	29	5	1	80	29	4	86	--	--	--	26	10	62	--	--	--	--	--	--
8/26/70	30	10	1	90	34	3	91	50	5	90	34	13	62	17.2	0.8	96	4.8	0	100
8/27/70	31	5	1	80	35	4	89	--	--	--	25	12	52	--	--	--	--	--	--
8/31/70	36	6	2	67	30	7	77	42	8	81	36	16	56	--	--	--	4.2	0.02	100
9/01/70	37	5	2	60	35	7	80	--	--	--	35	10	71	11.0	2.5	77	--	--	--
9/02/70	38	12	2	83	53	10	81	--	--	--	--	--	--	14.2	2.1	85	5.3	0.08	99
9/03/70	39	6	1	84	27	5	82	--	--	--	--	--	--	11.2	2.3	80	0.7	0.01	100
Average		8	3	67	36	5	85	57	9	83	37	13	65	12.3	1.8	85	2.7	0.02	99

through the carbon columns. These results indicated that the carbon column was not exhausted during the study, although it was approaching exhaustion. Projection of these data to breakthrough indicates that treatment of 14,000 gallons would exhaust carbon column one. This projected throughput corresponds to an exhaustion rate of 0.86 pounds of carbon/1,000 gallons of throughput. The exhaustion rate obtained through column testing was quite close to that obtained from the adsorption isotherm discussed above.

The results of the pilot plant indicated that carbon column four produces an effluent with a TOC concentration of approximately 10 mg/l for influent TOC concentrations varying from 25-61 mg/l. The performance of carbon column four in reducing organics, oil, and phenol is shown in Table 3. The average removal with respect to suspended solids, BOD_5, TOC, oil, and phenol were 62, 83, 65, 85, and 99 percent, respectively.

The pilot plant carbon columns were backwashed to maintain an acceptable pressure drop across the columns. However, samples for wastewater analysis were not taken during these backwashes.

Samples of exhausted carbon from the lead column were oven dried for three hours at 150°C prior to bench regeneration testing. The analytical results are presented below.

Sample	Iodine Numbers		
	Virgin Carbon	Exhausted Carbon	Regenerated Carbon
Top to Middle	900, min.	453	950
Middle to Bottom	900, min.	518	973

Based on these results, it was projected that spent carbon could be regenerated to its original adsorptive qualities based on iodine numbers using standard additions of steam and air/gas at 1750°F.

The average pilot plant influent and effluent values are summarized below.

Activated Carbon Adsorption Pilot Plant Average Performance

Oil			Suspended Solids			BOD_5		
mg/l		Percent	mg/l		Percent	mg/l		Percent
Inf	Eff	Removal	Inf	Eff	Removal	Inf	Eff	Removal
12.3	1.8	85	8	3	62	57	9	83

Activated Carbon Adsorption Pilot Plant Average Performance (cont'd)

TOC			Phenol		
mg/l		Percent	mg/l		Percent
Inf	Eff	Removal	Inf	Eff	Removal
37	13	65	2.7	0.02	99

SOLIDS DEWATERING STUDIES

Thickening of filter backwash and of a composite of filter backwash, API Separator bottoms, and emulsion treater bottoms were investigated. Laboratory tests indicated that the filter backwash could be thickened to 2 percent solids at a solids loading of 23 lb/sq. ft./day, and that the composite sludge could be thickened to a solids concentration of 1.7 to 3.0 percent at solids loadings of 23 to 62 lb/sq. ft./day.

Bench scale centrifugation of thickened sludge was investigated. The bench scale evaluation of centrifuge operation consisted of heating the sludge to 180-200°F. and centrifuging in a solid bowl centrifuge. The effluent from the solid bowl centrifuge was treated in a disc centrifuge to separate oil from the water layer.

SECTION 6

CONSTRUCTION AND START-UP

The following schedule was maintained for this project:

Engineering Complete	March 01, 1972
Bids Received	April 10, 1972
Contract Issued	May 01, 1972
Start of Construction	May 04, 1972
Completion of Construction	February 25, 1973
Start-up Commenced	February 26, 1973

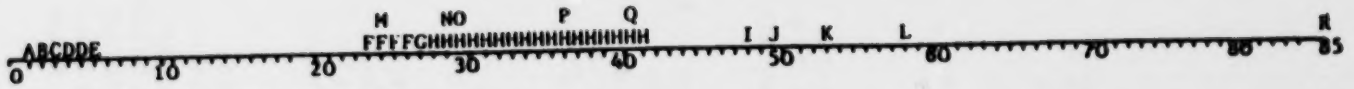
Start-up of the filters, adsorbers, and solids dewatering facilities followed a predetermined sequence. A summary of the start-up period is reported below. This summary of time sequential activities during start-up is shown in Figure 4.

Start-up of the dual media filters commenced on day one with a total flow of 400 gpm. Problems were encountered with the automatic butterfly valves, and the filters were shut down on day two for repairs. The filters were placed on line again on day three. Control adjustments were made on days four and five, and the flow was increased to 900 gpm on day six.

The adsorbers and the carbon storage tanks were topped out with carbon, using the carbon blow pot on days 23-26. At this time some irregularities were observed on the skirts of the adsorber vessels. During the next 14 days, 4 x 4 angles were welded to the skirts for additional structural support. Wastewater was reintroduced to the adsorber on day 48 with a total flow of 600 gpm. The influent rate was increased to 900 gpm on day 50.

On day 53, the flow to the filters and the adsorbers was increased to 1500 gpm and the units were switched over to automatic level control. As the level in the surge basin was lowered, the suspended solids which had settled out in the surge basin were scoured to the filter. The level in the surge basin had to be raised to reduce the suspended solids loading to the filters. The level in the surge basin was then slowly lowered at the rate of one inch per day.

The carbon regeneration furnace was initially started on



I. Dual Media Filter Startup

A. Filter startup commenced with flow of 400 gpm.
B. Filters shutdown for repair of automatic butterfly valves.
C. Filters placed on line again.
D. Filter control adjustments made.
E. Filter flow increased to 900 gpm.
K. Influent increased to design rate of 1500 gpm and units switched to automatic level control.

II. Carbon Adsorber Startup

F. Adsorbers and Carbon Storage Tanks topped out with carbon.
G. Irregularities observed in skirt of Adsorber vessels.
H. Additional structural support added to Adsorbed skirts.
I. Influent reintroduced to Adsorbers at 600 gpm.
J. Influent rate increased to 900 gpm.
K. Influent increased to design rate of 1500 gpm and units switched to automatic level control.

III. Carbon Regeneration Furnace Startup

L. Carbon Regeneration at design rate of 120 Pounds/Hour.

IV. Solids Dewatering System Startup

M. Filter backwash water introduced into Dewatering System.
N. Solid bowl centrifuge started.
O. Disc centrifuge started.
P. API Separator sludge introduced into Dewatering System.
Q. API Separator sludge discontinued due to plugging problems.
R. Comminuter installed and sludge transfer from API Separator resumed.

Figure 4. Summary of time sequential activities during startup of Filtration/Adsorption System.

day 26. Continuous regeneration of carbon at the design rate of 120#/hr. began on day 58. The solids dewatering system was started up on day 24 with the backwash water from the filters. The solid bowl centrifuge was started on day 29 and the disc centrifuge on day 30. On day 37 API Separator sludge was pumped to the sludge blending tank at the head of the solids dewatering system where it was passed through a one-quarter inch screen before entering the tank. This screen became clogged after one minute. Also, small sticks and other material which passed through the screen or were in the filter backwash water became clogged in the one-quarter inch pump impellers down stream of the sludge blending tank. It became necessary to discontinue the pumping of the API Separator sludge and the emulsion treater bottoms to the solids dewatering system, due to plugging of the screen and pump impellers.

A comminuter was installed on the solids collection tank outlet by day 85 and sludge transfer from the Separator was resumed.

SECTION 7

WASTEWATER TREATMENT SYSTEM DESIGN

A schematic flow diagram of the filtration/adsorption/regeneration/centrifugation wastewater treatment system is presented in Figure 5.

A discussion of the design of each individual system follows:

DUAL MEDIA FILTRATION DESIGN

Three parallel dual media filters were designed to remove oil and suspended solids from the API Separator effluent. Design removals were those achieved during pilot operation. An intermediate basin was included in the design to control flow surges and equalize influent overloads. Figure 6 is a cross section view of one of the dual media filters.

Each filter is a carbon steel vessel 10 feet in diameter by eighteen feet six inches overall height and is epoxy lined. Flow enters the bottom of the vessel and rises vertically through a 10 inch pipe in the center of the filter. A rated flow of 1000 gpm per filter corresponds to a superficial hydraulic loading of 12.8 gpm/ft.^{2}.

Flow to the filter system is controlled by a level controller which maintains a constant level in the intermediate surge basin. The effluent flow from each filter is sensed by individual flow indicators. The flows are summed and equally divided among the three filters by throttling each filter's effluent control valve.

Under normal filtering conditions the vessel is full of water to the vent connection on top. The water flow is down through the filter media of 2.5 feet of anthracite and 4.5 feet of sand; through the support gravel; and through the nozzles which are inserted in the supporting tube sheet. The water beneath the tube sheet flows out through the outlet connection to a 30,000 gallon filtered water holding tank.

Removal of suspended solids and oil trapped by the filters is accomplished by backwashing with water stored in the filtered water holding tank.

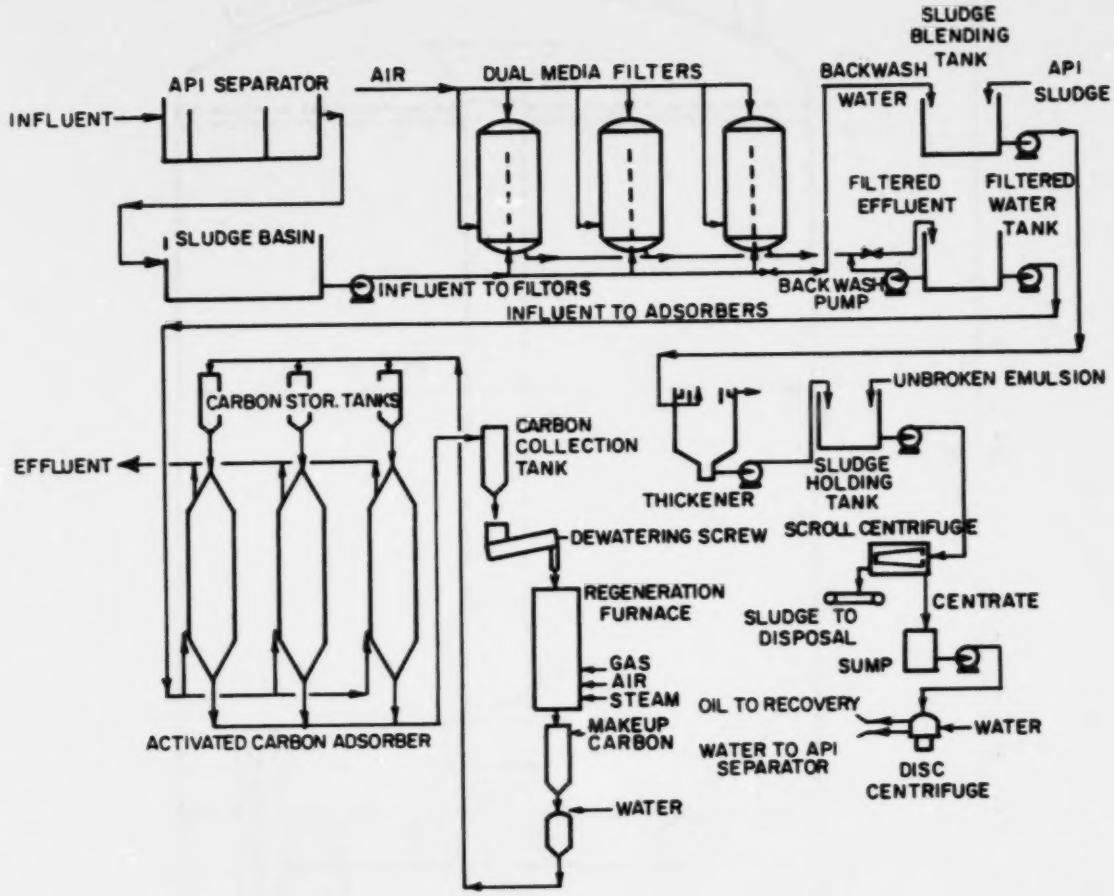


Figure 5. Wastewater Treatment Plant schematic flow diagram.

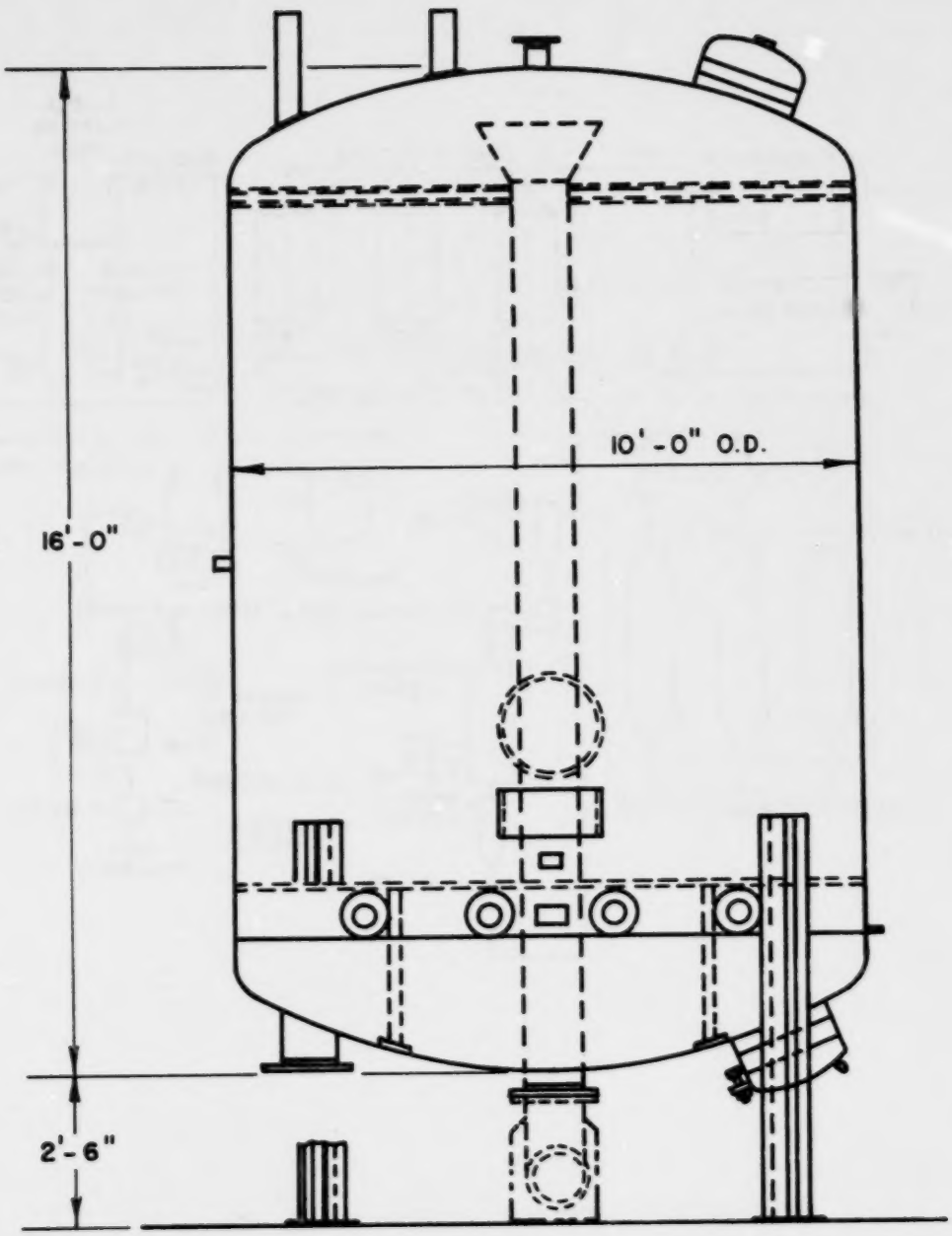


Figure 6. Dual media filter cross section

The initial step in backwash is to remove the water remaining in the filter by applying air pressure to the top of the vessel. This allows an up-flow air and water scour to follow and effectively remove adhering suspended solids and oil from the sand and anthracite particles. Scour rates are 7.1 gpm/ft.^{2} and 7.1 SCFM/ft.^{2}. As the scour water reaches the top of the vessel, the air is shut off and the water rate increases to 25.1 gpm/ft.^{2}, thereby flushing the filter of trapped suspended solids and oil. The backwash water overflows into the center standpipe and is directed to a 30,000 gallon sludge blending tank.

The backwash cycle is automatically operated by a Programmed Timer which can be initiated by an interval timer, high differential pressure, or manually by pushbutton. The three filter system is designed to allow only one filter to backwash at any one time. The filters will automatically backwash in numerical sequence. Although the filters were designed to operate with one off-line, the mode of operation is to have an individual filter off-line only during its backwash cycle. If the level in the filtered water holding tank is low or the level in the sludge blending tank is high, the backwash cycle cannot proceed and an alarm is sounded.

Table 4 summarizes the dual media filter design data.

ACTIVATED CARBON ADSORPTION DESIGN

Three parallel activated carbon adsorbers were designed to remove soluble organic matter from the filter effluent at a maximum flow rate of 2000 gpm. Design removals were those obtained during pilot operation. Figure 7 is a cross section view of one of the carbon adsorbers.

Each adsorber is a carbon steel vessel 10 feet in diameter by 65 feet overall height and is lined with 12-15 mils of Plastite. The adsorbers each contain 92,000 pounds of granular activated carbon in a bed depth of 45 feet. An additional 8000 pounds of carbon occupies the upper and lower end cone areas. The upper and lower cone angles are 90 and 46 degrees respectively, based on the angle of repose of granular activated carbon immersed in water.

Flow to the three adsorbers is controlled by the level in the filtered water holding tank, which acts as a feed surge basin. The influent to each adsorber is distributed through a circumferential manifold located just above the lower cone section. The flow is directed downward under an internal cone, then upward through a 3-foot diameter opening in the internal cone. A design flow to each adsorber of 667 gpm corresponds to an empty bed contact time of 40 minutes.

TABLE 4

Dual Media Filter Design Data

Filter Media		
Anthracite		
	Depth	2.5 FT
	Volume	195 FT^3
	Particle Diameter	0.25 IN
Sand		
	Depth	4.5 FT_3
	Volume	350 FT^3
	Particle Diameter	1 mm
Gravel Support		
	Depth	1.25 FT
Rated Flow (Each of Three Filters)		1000 GPM
Filter Diameter		10 FT
Center Standpipe		10 IN_2
Filter Area		78 FT^2
Hydraulic Loading		12.8 GPM/FT^2
Liquid Capacity With Media Installed		4800 GAL
Maximum Allowable Pressure Drop		
Thru Media		6.5 PSI
Design Pressure		47.5 PSI
Backwash Interval		12 HOURS
Backwash Water Flow As Percent of Filtrate		1.3%
	Low Rate	550 GPM 7.1 GPM/FT^2
	High Rate	1960 GPM 25.1 GPM/FT^2
Backwash Air Flow		550 SCFM 7.1 $SCFM/FT^2$

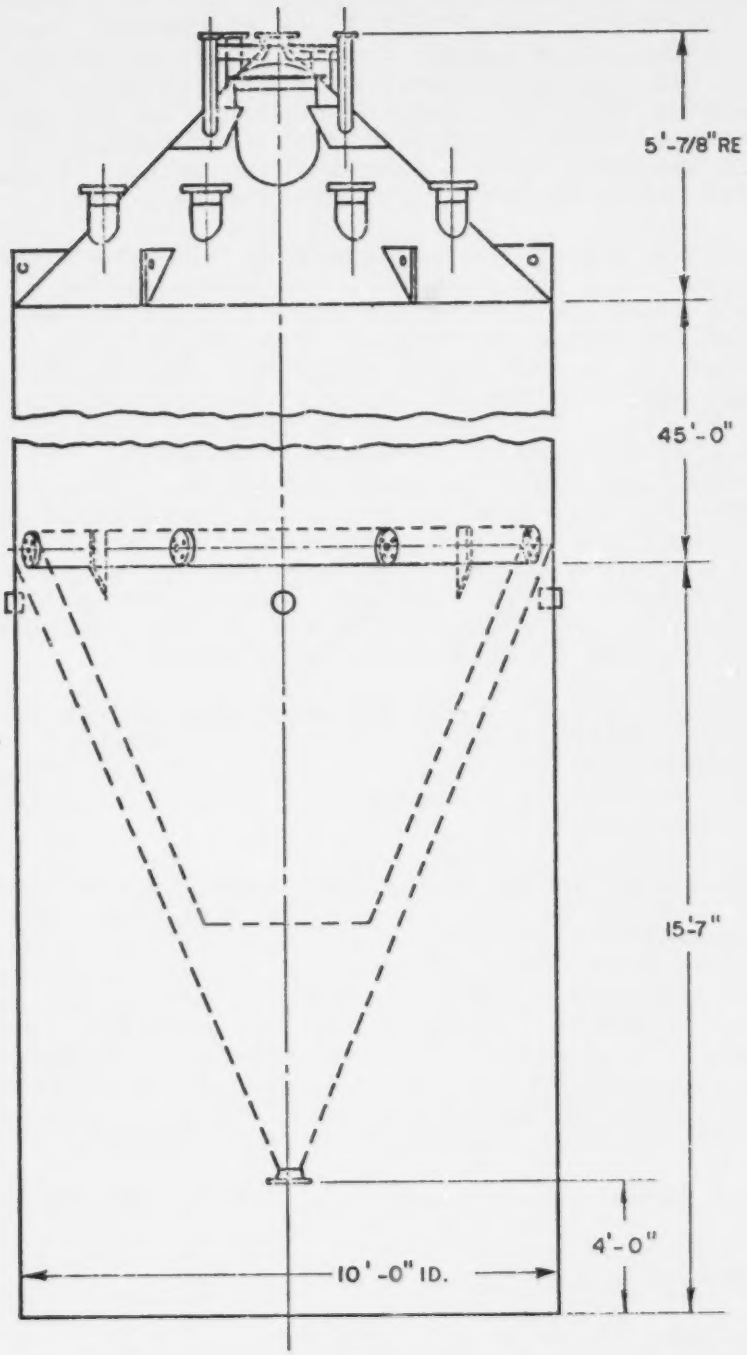


Figure 7. Carbon Adsorber cross section.

The upward flow through the packed bed at a superficial hydraulic loading of 8.5 GPM/ft.^{2}, is discharged through eight internal septums which extend vertically from the upper cone. The septums are stainless steel well screens which retain the 1.5 mm diameter activated carbon particles in the adsorber. Filtered service water is provided at each septum for backflushing, should plugging due to carbon fines occur. Figure 8 is a cross section view of one of the effluent septums.

Continuous adsorption is dependent upon the removal of exhausted carbon from the adsorbers and the addition of regenerated carbon. One thousand pounds per day of spent carbon is pulsed from each of the three adsorbers. This equates to 1.1 percent of the total bed of an individual adsorber. During the pulse period, which occurs for each vessel every 24 hours, the adsorber is taken out-of-service. The hydrostatic pressure available at the lower cone apex is used to transport the carbon slurry to a flooded collection tank. A pulse period of 1.4 seconds allows the desired 1000 pounds of carbon to be transferred under velocities of 5 feet per second. Transfer lines are 4-inch schedule 40 carbon steel with schedule 80 long radius sweeps. Ball valves are used in carbon slurry service. During this pulse period, regenerated carbon is added to the top of the adsorber from a carbon storage tank located above each vessel.

As the ball valve at the adsorber apex closes to stop spent carbon flow, filtered service water is introduced to flush the line, thereby preventing carbon bridging and corrosion. Freezing problems are avoided by draining the transfer line following completion of the water flush.

A cone bottom carbon collection tank receives the spent carbon and acts as the regeneration furnace feed tank. A ball valve at the apex of the collection tank pulses carbon for 8 seconds into a dewatering screw at two minute intervals. Filtered water is added at the apex to prevent carbon bridging, and is added to the dewatering screw to further wash the carbon of free oil which was "filtered out" in the adsorber. Overflows from the collection tank and dewatering screw are directed to a carbon settler from which the carbon is ejected into the dewatering screw, and the water overflows to be reprocessed.

Table 5 summarizes the activated carbon adsorption design data.

THERMAL REGENERATION DESIGN

A five foot diameter multiple hearth furnace was designed to thermally regenerate the spent carbon. The dewatered carbon enters the six hearth furnace through an 8-inch inlet for regeneration at a design rate of 125 pounds per hour. The regenera-

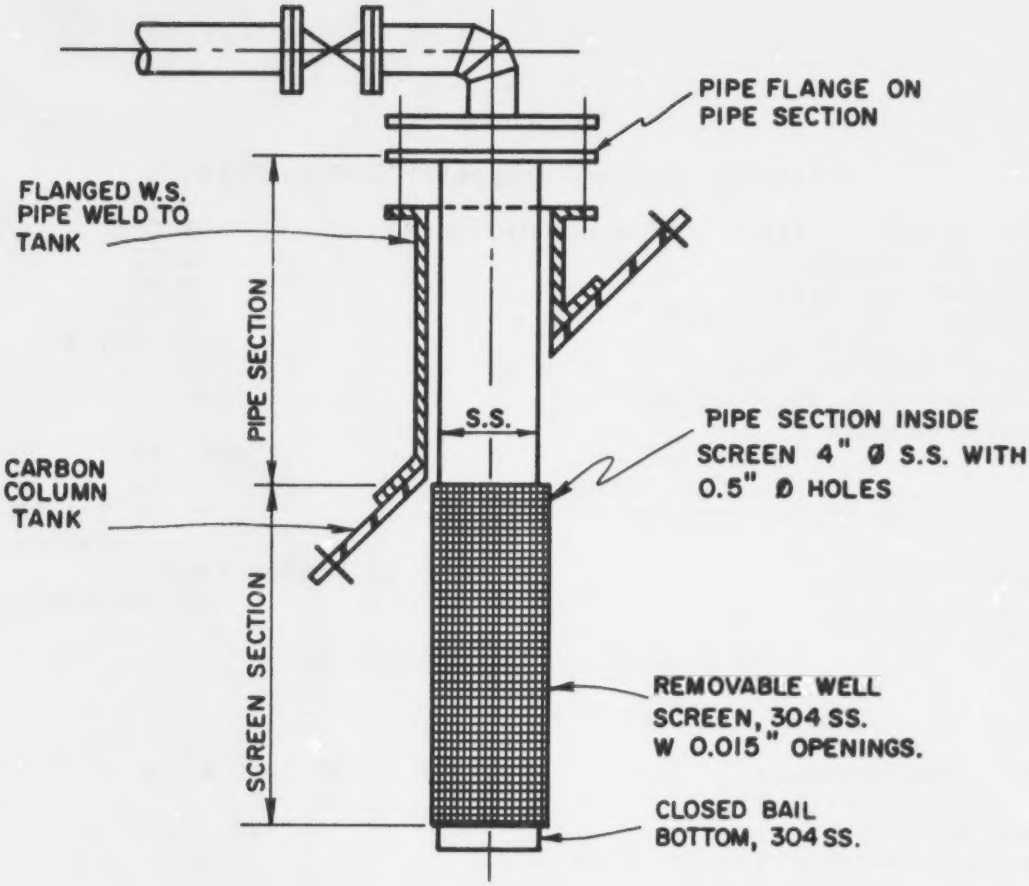


Figure 8. Carbon Adsorber effluent septum cross section.

TABLE 5

Activated Carbon Adsorption Design Data

Rated Flow (Each of Three Adsorbers)	667 GPM
Adsorber Diameter	10 FT
Adsorber Bed Depth	45 FT
Contact Time (Empty Bed)	40 MIN
Hydraulic Loading	8.5 GPM/FT^2
Design Inlet Pressure	60 PSI
Pressure Drop Thru Carbon	35 PSI
Carbon Inventory	
Carbon Bed	92,000 LB
Adsorber Total	100,000 LB
Theoretical Carbon Capacity	0.3 LB TOC/LB Carbon
Carbon Dosage	0.86 LB Carbon/1000 GAL Throughput

Activated Carbon Properties

Filtrasorb 300

Total Surface Area (N_2 BET Method)	950-1050 M^2/g
Bulk Density	26 LB/FT^3
Particle Density Wetted in Water	1.3-1.4 g/cc
Mean Particle Diameter	1.5-1.7 mm
Iodine Number, minimum	950
Ash	Max 8%
Moisture	Max 2%

tion furnace is capable of handling up to 250 pounds per hour. The carbon is moved downward through the fire brick lined hearths by cast iron rabble arms. In the first hearth, which is unfired but maintains a temperature of 1100°F, any remaining moisture is vaporized. Hearths four and six, numbered from the top, are tangentially fired by two burners using refinery fuel gas at rates of 188 and 68 CFH respectively, to maintain respective temperatures of 1725°F and 1750°F.

In an atmosphere controlled by addition of steam at a design rate of 125 pounds per hour, the adsorbed organics are volatilized and oxidized. To assure complete oxidation, all flue gases pass through an integral afterburner fired by refinery fuel gas and maintained at a temperature of 1350°F. Recirculation of shaft cooling air provides sufficient oxygen for combustion. Prior to emission to the atmosphere, the flue gases pass through a two-foot diameter, four plate, wet scrubber using filtered service water for gas cooling and particulate removal to 0.04 grains per standard cubic foot (dry).

Temperature indicator controllers maintain the desired temperature in the fired hearths. Furnace safety features include ultra-violet flame scanners and alarms which annunciate should the combustion air blower, induced draft fan, or the shaft cooling air fan fail. Abnormally high or low fuel gas pressure will cause the main gas safety valve to close, resulting in a flame-out at all burners.

Regenerated carbon is discharged from the furnace into a 12 cubic foot cone bottom quench tank flooded with filtered service water. Temperature reduction, the addition of make-up carbon, and the formation of a carbon slurry occur in the quench tank. As the carbon level in the quench tank increases, a rotating bindicator is stopped and a timed sequence is initiated to transfer the regenerated carbon to one of three 96 cubic foot carbon storage tanks located above each adsorber.

During the time controlled sequence, the carbon slurry flows by gravity into a 5 cubic foot blow case. Filtered service water is then introduced into the blow case to pressure the carbon at velocities of 5 feet per second through 2-inch transfer lines of schedule 40 carbon steel with schedule 80 long radius sweeps. The slurry transfer is followed by a water flush and an air drain to clear the line. In the event a high level is indicated by a storage tank bindicator, the carbon is automatically transferred to the next storage tank.

Carbon addition to the adsorbers from the storage tanks, which occurs during the pulsing of spent carbon from the bottom cone, is judged complete by a bindicator located in the upper cone. Should the bindicator indicate a low level, the adsorber may not be brought back into service.

An additional safety feature is an atmospheric vent from the top of the adsorber to its carbon collection tank. In the event a number of septums plug simultaneously, excess flow will be vented, and overflow the collection tank to the carbon settler. A pressure gage is located on the vent line to indicate such an occurrence.

Table 6 summarizes the regeneration furnace design date.

SOLIDS DEWATERING SYSTEM DESIGN

A solids handling system was designed to separate the sludge removed at the Filters, API Separator, and Emulsion Treater into an oil, water, and solid phase. Upon separation, the oil is recovered, the water is returned for reprocessing, and the solids are disposed at an offsite licensed sanitary landfill.

The three intermitent sludge streams noted above are mixed in a 30,000 gallon sludge blending tank and transferred to a 26,000 gallon circular thickener at a rate of 60 gpm. The design loading of 30 pounds per square foot per day results in an underflow concentration of 2.5 percent solids. The thickener underflow of 20 gallons per minute and overflow of 40 gallons per minute are directed to the sludge holding tank and API Separator respectively. Should an emulsion layer accumulate on the thickener, it is skimmed directly to the sludge holding tank. The sludge holding tank acts as a feed surge basin for the scroll centrifuge.

Feed to the scroll centrifuge, flowing at 20 gpm, passes through a double pipe heat exchanger which maintains an outlet temperature of 150°F. Operating at 2600 RPM, the scroll centrifuge discharges a stream of 50 percent solids, and an oil-water stream. The solids are carried by conveyor belt to a holding container to await disposal. The liquid centrate is directed to the disc centrifuge feed sump.

The disc centrifuge feed, at 20 gpm. passes through a double pipe heat exchanger which maintains an outlet temperature of 180°F. An additional 25 gpm of filtered service water also enters the disc machine at 180°F to establish a nozzle seal. Operating at 6350 RPM, the disc centrifuge discharges an oil stream for recovery, a water stream for reprocessing, and a solids-water stream, also for reprocessing through the API Separator.

THERMAL REGENERATION PERFORMANCE

Spent carbon regeneration has been achieved using a six-hearth furnace fired by refinery fuel gas. The maximum regeneration rate has been 250 pounds per day. The maximum steam addition rate for control of the furnace atmosphere has been 250 pounds per hour.

TABLE 6

Thermal Regeneration Design Data

Furnace	60" x 6 Hearth with Integral Afterburner
Regeneration Rate	125 LB/HR
Steam Addition Rate	125 LB/HR
Fuel	Refinery Fuel Gas
Fuel Rate	
Hearth 4	188 CFH
Hearth 6	68 CFH
Afterburner	310 CFH
Combustion Air Rate	
Hearth 4	5000 CFH
Hearth 6	1800 CFH
Afterburner	8120 CFH
Design Temperatures	
Hearth 4	1725°F.
Hearth 6	1750°F.
Afterburner	1250°F.

A 40 percent decrease in the adsorptive capacity of the regenerated carbon was observed following eighteen months of operation. Regenerated carbon iodine numbers in the range of 560-680 have been determined and show a decrease from virgin carbon iodine numbers, which are in the range of 950-1000. Regenerated carbon molasses numbers of 280 show an increase over the virgin carbon molasses number of 230. A decrease in micro pores and an increase in macro pores in the activated carbon are indicated by the above results.

Adsorption isotherms were prepared using both regenerated and virgin carbon and resulted in the following loadings at current influent concentrations.

Virgin Carbon	Loading	Regenerated Carbon
0.17	Pounds TOC/Pound Carbon	0.096
0.73	Pounds COD/Pound Carbon	0.35
0.04	Pounds Phenol/Pound Carbon	0.03

In order to achieve greater regenerated carbon adsorptive capacity, a revised regeneration furnace profile is currently under evaluation. Rather than the design gradual temperature increase through the furnace hearths, the revised profile maintains a temperature of 1200°F in number four hearth and increases to 1750°F in number five hearth. The purpose of this rapid transition is to pass through the coking range, thereby preventing plugging of the carbon micro pores.

Carbon regeneration, as measured by the carbon's apparent density and comparing the value with virgin carbon, has been determined to be a function of regeneration rate. With the carbon out interval set at 4.5 hours, i.e., a 3000 pounds per day regeneration rate, the regenerated carbon's density averaged 51.8g/100 cc. With the carbon-out interval set at 8 hours, i.e., a 1500 pounds per day regeneration rate, the regenerated carbon's density averaged 50.0g/100 cc. The spent carbon averaged 59.4 and 59.5 during these periods respectively.

A tar-like substance consisting of carbon fines and water has caused plugging of the flue gas transfer line and wet scrubber. This plugging results in a backpressure on the furnace which activates the automatic furnace shutdown mechanism. The scrubber trays and piping must then be removed for cleaning.

Two one-quarter inch lines were installed in the quench tank to provide filtered water to flush the regenerated carbon away from the furnace drop chute. Prior to this installation, regenerated carbon had backed up on the chute and into No. 6 hearth of the furnace, causing excessive metal temperatures.

Carbon losses have recently been six percent per regeneration cycle. Previous carbon losses were in the eight to ten percent range. These initial high loss values may be attributed to mechanical problems, carbon loss in the effluent, and carbon losses which occur when an adsorber was overpressured and vented to the atmosphere. The losses attributable to each of these factors has not been determined.

The percent of time that the regeneration furnace has been off line for maintenance has increased during each year of operation. During the first year of operation, the furnace was off line 6 percent of the time. The percent offline during the second year was 16. During the third year the furnace was off line 22 percent of the time.

At various times the furnace was shutdown in order to: (1) clean the scrubber and flue gas line of built-up tar; (2) clean the combustion air lines and gas lines and controls of corrosion products; (3) repair cracks in the flue gas line; (4) replace the furnace sand seal; (5) realign the ultraviolet flame sensors; or (6) other routine maintenance.

One particular furnace shutdown was caused by the inability to transfer regenerateratered carbon due to the number of leaks which developed in the two-inch carbon transfer lines. The leaks were the result of corrosion along the bottom section of the transfer line in the horizontal run between the blow case and vertical pipe section. The determination of pipe wall thickness along the remainder of the carbon steel pipe and bends revealed no other significant reduction in wall thickness. It is suspected that carbon lying in the horizontal line resulted in corrosion of the carbon steel wall.

A number of shutdowns were caused by failure of the flue gas line from stress corrosion cracking probably the result of chloride attack.

The corrosion in the combustion air lines and controls was determined to be the result of carbon fines entering the suction of the combustion air blower. The source of the fines was the addition of makeup carbon to the quench tank which was located adjacent to the combustion air blower. A filter and protective cover were added to the blower to precent the entrance of carbon fines and excess water.

SOLIDS DEWATERING SYSTEM PERFORMANCE

The scroll centrifuge has demonstrated average removals of 30 percent for suspended solids. The disc centrifuge has demonstrated average removals of 91 percent for oil. Performance data for the scroll and disc centrifuges follows:

Centrifuge performance

	Percent Solids	Percent Oil
Scroll Centrifuge Feed	11.6	2.7
Scroll Centrifuge Centrate	0.7	5.9
Scroll Centrifuge Sludge	33.6	5.5
Disc Centrifuge Water Discharge	0.1	0.3
Disc Centrifuge Solids Discharge	0.4	0.3

Scroll Centrifuge 13 gpm
Disc Centrifuge 43 gpm (including 30 gpm utility water)

A major problem associated with the solids dewatering system is the recycle of solids and oil to the API Separator influent. This recycle has increased the solids and oil loading and emulsion volume on all downstream treatment units. This increase has in turn overloaded the solids dewatering units and has resulted in the "back-up" of sludge and emulsion in the API Separator.

The average individual contributions to the solids system based on two one-week surveys are reported below:

		#/Day	
Source	Gal/Day	Oil	Solids
Emulsion Treater	21,300	46,150	5,160
Filter Backwash	45,000	480	520
API Separator	2,000	630	720
Total	68,300	47,260	6,400

During this period 27,500 pounds of oil and 5,700 pounds per day of solids were recycled to the separator influent. The sources of this recycle are the thickener overflow and disc centrifuge.

Sludges from the API Separator and Emulsion Treater caused severe plugging problems in the solids dewatering system. The inlet screens at the sludge blending tank, two inch transfer lines, pump impellers, and the double pipe heat exchangers experienced plugging due to the debris contained in these sludges. The problem has been corrected by the installation of a comminutor.

In order to improve operation of the sludge thickener, and reduce recycle to the separator, unbroken emulsion from the Emulsion Treater currently bypasses the sludge thickener and is transferred directly to the sludge holding tank. Rather than using the design method of continuously transferring to the

sludge holding tank, thickener bottoms are pumped directly to the scroll centrifuge for 30 minutes during each shift.

The wear plates on the sludge pumps had to be renewed after two years of operation. The abrasive carbon fines are suspected to be the cause of the excess wearing.

In order to maintain a minimum sludge velocity of five feet per second, the scroll centrifuge feed piping has been reduced from two inch to one and a quarter inch diameter. Following the installation, an improvement in both the scroll and disc centrifuge operation has been observed. The solids concentration of the scroll cake has increased and the disc centrifuge has maintained a longer run time between shutdowns.

An internal inspection of the scroll centrifuge, following nine months of operation, revealed that the hard surface coating on the internal flights had worn away exposing the stainless steel. One-half inch thick, layered, deposits of carbon fines and grit were found inside the scroll. The scroll was returned to the manufacturer for resurfacing, and a spare scroll was obtained.

The internal flights required a second resurfacing after being in operation for 24 months. The increase in on line time may be attributed to a decrease in the recycle of carbon fines to the solids handling system.

A strainer with continuous backwash was installed upstream of the disc centrifuge to remove solids which escaped the scroll centrifuge. Carbon fines not removed by the scroll centrifuge have on occasion eroded the screen of the strainer.

The disc centrifuge sealing water has been flow controlled at 25 gpm to prevent the flushing of bearing grease which occurred previously, causing a bearing burnout and resulting in an extended shutdown period. The maximum period that the disc centrifuge has been on stream without plugging has been 34 days. The downtime required for removal of the disc stack and cleaning can extend to three days. During this downtime, centrate from the scroll centrifuge is recycled to the API Separator. Visual inspection of the disc stack has revealed a coating of carbon fines.

The factors discussed above have contributed to increasing downtime for the centrifuge system. During the first year of operation the centrifuge system was off-line 29 percent of the time. During the second year the off-line time increased to 37 percent and increased to 42 percent during the third year.

The rate of oil recovered from the disc centrifuge is a function of the operation of the API Separator, the Emulsion

Treater, and the sludge handling facilities. Oil recovery has approached 20 gpm on occasion.

Construction of inlet sludge pumping stations at the API Separator has permitted removal of sludge from the separator inlets. This has resulted in improved separator operation, and a reduction in the volume of sludge directed to the solids handling facilities.

SECTION 8

WASTEWATER TREATMENT SYSTEM PERFORMANCE

DUAL MEDIA FILTRATION PERFORMANCE

The dual media filters demonstrated average removals of 68 percent for suspended solids and 75 percent for oil during the first year of operation. These averages are taken from the data reported in Tables 7-10. Although individual operating period characteristics have varied, performance of the filters remained uniform during the first two years of operation. The filter's removal efficiency decreased as media was gradually lost from the filter vessels. This loss in efficiency is evidenced by the data reported in Table 11 for operations during the third year after start-up. For the period January, 1975-December, 1975 the mean removal for suspended solids was 59 percent and 36 percent for oil. Table 11 also reports the filter influent and effluent mean and maximum concentrations for suspended solids, oil, TOC, COD, phenol, and sulfide for the period January-December, 1975. During this operating period the mean flow to the filters was 1939 gpm with a maximum of 2290 gpm.

Additional data on the filter operations is reported in the appendix. Histograms are plotted for the frequency of occurrance of suspended solids, oil, TOC, COD, phenol and sulfide concentrations for the year January-December, 1975. Also reported is the mean, standard deviation, maximum and minimum for each parameter.

Table 12 presents an analysis of filter backwash water. Based on this data, the backwash high rate flush duration was increased to 7 minutes and the flow rate was increased to greater than 2000 gpm. The total backwash duration averages 20 minutes and is dependent upon the time required for the pressurized removal of water remaining in the filter. The time required is a function of the differential pressure across the filter when a backwash is initiated. The backwash interval is set at 4 hours, i.e., each individual filter is backwashed every 12 hours.

Although the design included a differential pressure override to initiate backwash, this option has not been used since the maximum differential pressure reached during the above backwash interval has been 3 psi.

TABLE 7

Wastewater Analyses

Period 1

April 1973

- Initial Operation
- Virgin Carbon
- Foul Condensate Not Included

		Concentration, ppm			Percent Removal	
Parameter		Filter Influent	Adsorber Influent	Adsorber Effluent	Filter	Adsorber
Suspended Solids	Avg	70	16	11	77	31
	Max	95	29	14		
BOD_5	Avg	76	46	40	39	13
	Max	-	-	-		
COD	Avg	386	216	63	44	71
	Max	478	248	77		
Oil	Avg	21	10	0.34	52	97
	Max	34	19	0.72		
Phenol	Avg	21	20	0.023	5	99.9
	Max	21	20	0.028		
Sulfide	Avg	-	0.19	13.8	-	Increase
	Max	-	0.35	18		
Ammonia	Avg	-	15.2	19	-	Increase
	Max	-	16.2	20		
Flow	Avg	1485	1375			
	Max	1520	1710			

TABLE 8

Wastewater Analyses

Period II

July 1973

- Early Regeneration
- Carbon Bed Not Turned Over
- Foul Condensate Included

		Concentration, ppm			Percent Removal	
Parameter		Filter Influent	Adsorber Influent	Adsorber Effluent	Filter	Adsorber
Suspended Solids	Avg	88	34	21	61	38
	Max	115	52	67		
BOD_5	Avg	75	78	72	-	8
	Max	-	-	-		
COD	Avg	388	296	133	24	55
	Max	522	361	184		
Oil	Avg	74	25	7.7	66	69
	Max	143	66	28		
Phenol	Avg	14	12	4.5	14	63
	Max	19	22	8.4		
Flow	Avg	1850	1405			
	Max	1900	1505			

TABLE 9

Wastewater Analyses

Period III

October 1973

- Carbon Bed Turned Over
- Maximum Regeneration Rate
- Foul Condensate Included

		Concentration, ppm			Percent Removal	
Parameter		Filter Influent	Adsorber Influent	Adsorber Effluent	Filter	Adsorber
Suspended Solids	Avg	67	20	16	70	20
	Max	126	34	43		
COD	Avg	415	322	242	22	25
	Max	660	400	300		
Oil	Avg	67	10.4	2.1	84	80
	Max	114	16.3	3.2		
Phenol	Avg	32.5	32	12.9	-	60
	Max	34.5	33.5	20.0		
Sulfide	Avg	-	7.5	37	-	Increase
	Max	-	16.8	11.2		
Ammonia	Avg	-	87	93	-	Increase
	Max	-	103	110		
Flow	Avg	1855	1420			
	Max	1960	1600			

TABLE 10

Wastewater Analyses

Period IV

February 1974

- Effluent Septums Bent
- Foul Condensate Not Included
- One Year of Operation

		Concentration, ppm			Percent Removal	
Parameter		Filter Influent	Adsorber Influent	Adsorber Effluent	Filter	Adsorber
Suspended Solids	Avg	64	16	55	75	Increase
	Max	74	29	59.0		
BOD_5	Avg	88	55	65	36	Increase
	Max	-	-	-		
COD	Avg	408	301	253	25	13
	Max	500	390	330		
Oil	Avg	78	13	8	83	38
	Max	125	22	16		
Phenol	Avg	2.0	1.9	0.7	-	63
	Max	3.8	3.7	0.8		
Sulfide	Avg	-	0.6	13	-	Increase
	Max	-	0.9	25		
Ammonia	Avg	-	12.3	13.0	-	Increase
	Max	-	13.5	15.0		
Flow	Avg	1770	1030			
	Max	1970	1175			

TABLE 11

Wastewater Analyses

Period V

January 1, 1975-December 31, 1975

		Concentration, ppm			Percent Removal	
Parameter		Filter Influent	Adsorber Influent	Adsorber Effluent	Filter	Adsorber
Suspended Solids	Mean	116	48	45	59	6
	Max	507	396	999		
TOC	Mean	162	122	71	25	42
	Max	479	362	384		
COD	Mean	475	331	195	31	40
	Max	940	880	1012		
OIL	Mean	47	30	14	36	53
	Max	168	110	99		
Phenol	Mean	14.2	13.5	15.1	5	Increase
	Max	85	80	76		
Sulfide	Mean	-	2.3	14.3	-	Increase
	Max	-	32	71		
Flow, GPM	Mean	1939	731			
	Max	2290	1252			

TABLE 12

Analyses of Sand Filter Backwash Water

Time After Initiating High Rate (1960 GPM) Backwash Mins.	Suspended Solids ppm
0	87
1	19,370
2	9,010
3	4,160
4	3,010
5	850
6	300

TABLE 13

Analyses of Water Removed from Sand Filters by Pressure Prior to Backwash

Time After Air Pressurization Minutes	Suspended Solids ppm
0	13
1	42
2	296
3	196
4	156
5	77
6	91
7	2,010
8	5,140
9	4,900

The factor limiting the backwash interval has been the capacity of the sludge blending tank which receives backwash water.

An operating problem encountered was a decrease in effluent quality due to a backwash cycle. The two filters remaining on line experience a "shock" as the individual flow rates increase to include that portion of flow previously handled by the third filter. The decrease in effluent quality was due to the expelling of suspended solids and oil trapped in the filter, and the decrease in removal at the increased hydraulic loading. An increase in effluent quality was obtained by setting the controls to maintain the established flow rate to each on-line filter during a backwash cycle.

A decrease in effluent quality was also observed during the removal of water remaining in the filter by applying air pressure prior to backwash. Table 13 presents an analysis of this water and indicates that suspended solids are removed from the filter and contaminate the effluent water stored in the filtered water holding tank.

A draindown system to return water pressurized from the filters during the first stage of backwash to the API Separator was installed to increase the quality of filtered water. The suspended solids removal efficiency increased from a range of 27-68 percent to a range of 45-76 percent following activation of the draindown return to the API Separator. The oil removal efficiency increased from a range of 47-52 percent to a range of 68-86 percent. Evaluation of selective similar influent grab sample values for oil, occuring on days with approximately equivalent flow rates led to the above reported percent reductions. The reported suspended solids values were derived from a 24 hour composite collected on the day of the reported oil values. The data base for the above evaluation is presented in Table 14.

Over a period of 18 months of operation, 13 inches of anthrafilt was lost from the filter vessels. During the subsequent 18 months of operation, an additional 6 inches was lost from the vessels. Also, over this 36-month operating period, 12 inches of sand was lost from the vessels.

ACTIVATED CARBON ADSORBER PERFORMANCE

The adsorbers demonstrated varying performance over the one year operating period from March, 1973, through February, 1974, as a function of individual operating period characteristics.

Table 7 presents influent and effluent concentrations observed during the Period 1 when the adsorbers were in initial operation with minimum carbon bed pulsing and virgin carbon. The foul

TABLE 14

Dual Media Filters Draindown Data

OIL

Draindown to Filtered Water Tank

Flow Dual Media Filter Influent		Effleunt	%Reduction
GPM	ppm	ppm	
1502	132	69.7	47%
1757	111.7	57.7	48%
1405	71.8	40	44%
1530	53.1	25.2	52%

Draindown to API Separator

Flow Dual Media Filter Influent		Effluent	%Reduction
GPM	ppm	ppm	
1470	135.5	25.3	81%
1684	123.1	39.6	68%
1467	75.9	13.4	82%
1539	57.4	8.2	86%

SUSPENDED SOLIDS

Draindown to Filtered Water Tank

Flow Dual Media Filter Influent		Effluent	%Reduction
GPM	ppm	ppm	
1502	91	66	27%
1757	97	67	31%
1405	41	23	44%
1530	38	12	68%

Draindown to API Separator

Flow Dual Media Filter Influent		Effluent	%Reduction
GPM	ppm	ppm	
1470	98	54	45%
1684	73	28	62%
1467	55	13	76%
1539	53	13	75%

condensate stream from the Fluid Catalytic Cracker was not included with the process wastewater. With minimum carbon bed pulsing, an essentially static bed resulted in removal of oil to 0.34 ppm as measured by Freon extraction and infra-red absorption. This removal was not observed following full scale regeneration and normal bed movement. Phenol removal approached 100 percent during this period due to the low influent loading and a low spent carbon wave front. The increase in sulfide concentration from an influent of 0.19 ppm to an effluent of 13.8 ppm, is sulfide production occurring during initial operation.

The influent concentration from organic sulfur compounds was determined to be 0.01 ppm. The sulfide present in the effluent created an effluent odor problem and further investigation indicated the presence of butyl mercaptan, thiophene, and dimethyl sulfide.

The production of sulfide was not observed during a period when the refinery was shut down for maintenance and modernization. Total organic carbon results on the adsorber influent and effluent indicated that organic material was being removed during the shutdown period. The production of sulfide was also not observed during operation of a trial pH adjustment system at the API Separator. With the API Separator inlet pH maintained at 6.5-7.0, an increase in sulfide concentration did not occur across the carbon adsorbers. It was concluded that it is not just the presence of bacterial action which results in sulfide production, but rather it is a function of the influent content. This conclusion was not pursued further.

The difference between the reported filter and adsorber flow rates is due to the utilization of filtered water as unit service water and as backwash water.

The influent and effluent concentrations observed during Period 2 are presented in Table 8. The foul condensate stream from the Fluid Catalytic Cracker was included with the process wastewater during this period.

The increase in effluent phenol concentration observed during this period may be attributed to the introduction of 100 gpm of stripped foul condensate containing an average 300 ppm phenol. The adsorptive capacity of the carbon for phenol was 0.03 pounds of phenol per pound of carbon, based on an adsorption isotherm. The introduction of foul condensate therefore resulted in an influent phenol overload.

During this period the spent carbon wave front moved upward through the carbon bed. In order to achieve increased removals, the carbon regeneration rate was increased to 250 pounds per hour, thereby providing additional adsorptive capacity. It was observed that during the period in which the spent carbon wave front

had moved upward in the adsorber, an increase in effluent phenol concentrations occurred at low adsorber influent concentrations. This was a result of the adsorbed phenol achieving equilibrium with the phenol in solution in the wastewater. Again, lowering the spent carbon wave front will provide additional adsorptive capacity and eliminate the occurrence of this phenomenon.

Table 9 presents influent and effluent concentrations during Period 3 when the carbon bed had turned over, the regeneration rate was at the maximum 250 pounds per hour, and the foul condensate stream was included with the process wastewater.

The increase in effluent COD concentrations observed during Periods 3 and 4 may be attributed to a change in influent wastewater characteristics. During the design stages, the adsorption isotherm prepared using virgin carbons indicated a theoretical loading of 0.3 pounds TOC per pound carbon. However, current isotherms using virgin carbon resulted in an average loading of 0.17 pounds TOC per pound carbon, thus indicating a change in influent characteristics. This loading corresponds to an exhaustion rate in the range of 2.9-6.3 pounds of carbon per 1000 gallons of throughput, as compared to the exhaustion rate of one pound of carbon per 1000 gallons of throughput predicted by the initial isotherm.

The influent and effluent concentrations observed during Period 4 are presented in Table 10.

The Wastewater Treatment Plant had been in operation for a one-year period. The adsorbers operated during this period with flow rates ranging 630 to 1175 gpm. It is suspected that excessive carbon fines in the carbon beds and fines plugging the effluent septums were causing the flow restriction.

Table 11 presents adsorber influent and effluent concentrations during Period 5, January-December, 1975. The adsorbers operated during this period with a mean flow rate of 731 gpm. Carbon fines were continuing to plug the carbon beds and effluent septums. During this operating period, the mean removal of COD was 40 percent, for TOC the removal was 42 percent and for oil the mean removal was 53 percent.

Additional data on the adsorber operations is reported in the appendix. Histograms are plotted for the frequency of occurrance of suspended solids, oil, TOC, COD, phenol and sulfide concentrations for the year January-December, 1975. Also reported is the mean, standard deviation, maximum and minimum for each parameter.

The excessive suspended solids discharge reported in Tables 10 and 11 was due to holes in the effluent septums. The effluent

septums were bent inward creating openings in the screens. In order to correct the septum bending problem, new septums were installed. The new septum design included an internal four inch diameter stainless steel sleeve with one-half inch bored holes to provide structural strength to the external 0.015 inch slotted stainless steel screen. A deformation of the upper adsorber cone has been observed at the location of various septums. It is suspected that this deformation is the result of force transmitted to the three-eighth inch thick carbon steel cone when a septum was bent by the rising carbon bed.

Carbon fines had been removed from the system by flushing the carbon adsorbers. The flush is accomplished by closing the adsorber effluent valves and introducing 150 gpm into the base of the vessel and allowing the bed to "fluff" into carbon storage tank above the adsorber. The flush water containing the carbon fines overflows the carbon storage tank to the carbon settler, from which the water is directed to the API Separator and the settled carbon fines are transferred to the regeneration furnace where they are combusted.

The adsorbers were flushed after every "carbon out" cycle. This flush, together with a reduced regeneration rate, resulted in a decrease in carbon fines discharged with the effluent water when full flow is gradually returned to the adsorbers. The flushing also resulted in a significant decrease in adsorber effluent oil concentrations. However, flushing the adsorbers to remove carbon fines were discontinued. It was suspected that this flushing was generating additional fines, was contributing to the septum bending problem, and was causing gaps in the carbon bed within the adsorber.

Eight new septums were installed in each of the three carbon adsorbers. Prior to installation of the new septums, the total adsorber flow rate ranged from 500-1015 gpm, while the effluent suspended solids concentration ranged from 2-415 ppm. Following installation of the new septums, the total adsorber flow rate increased in range from 1035-1535 gpm, while the effluent suspended solids concentration decreased in range from 9-32 ppm.

Attempts were made during a "carbon out" cycle to backflush the effluent septums while the carbon was at low level, and the adsorber influent flow was shut off. This attempt did not result in an increase of flow through the adsorber.

An activated carbon pilot plant was operated in parallel with the full scale adsorbers in an effort to determine the cause of the flow restrictions observed in the adsorbers.

The pilot plant was monitored by taking flow and pressure readings through four 4 inch diameter by 4 foot length glass columns. The pressure gauges were located to permit the deter-

mination of pressure drop across the carbon bed or across the inlet or outlet septum. Regenerated carbon was used to fill the columns.

Initially, 50 psi inlet pressure was required to achieve the design flow of 2.9 gpm. The pressure drop was occurring at the outlet septum of each column. No readable pressure drop was occurring across the columns per se, only across the septum.

After one hour of operation, the design flow could not be maintained. Visual observation indicated that carbon fines were concentrating at the top of each column around the outlet septum.

The columns were hydraulically defined and the pilot plant operation was continued. This time only 5 psi feed pressure was required to achieve the design flow of 2.9 gpm.

The pressure drops were about equally divided over the four towers and outlet septum. The pilot plant was operated continuously for a two week period. The feed pressure had to be increased to 20 psig to maintain the design flow rate. During the first week, obvious visual evidence of biological growth was present in the carbon beds. However, the pressure drop through the four beds amounted to a total of less than 0.5 psi. The remainder of the pressure drop occurred at the outlet septum. The pilot plant was operated for another two weeks and the same effect was observed, except that a feed pressure of 40 psig was required to maintain the target flow rate.

At the end of the one month period, the towers were disassembled. The outlet septums were caked with a gritty black powder which x-ray identified as carbon fines. The carbon beds were coated with a slimy brown-black mass which from visual appearance and unique odor was bacterial. A bacterial plate count of the material around the outlet septum showed 15×10^4 organisms/gm, while the count through the bed was greater than 10^8 organisms/gpm.

In conclusion, bacterial growth was not the cause of the flow reduction observed in the pilot plant. Fines pluggage of the outlet septum appears to be the main cause of the flow reduction observed in the carbon columns.

The septum backwash transfer piping has been increased to three inch, with a two inch manifold at each adsorber to provide adequate flow to remove carbon fines that plug the septums and restrict flow. The design provides for a backflush of 90 gpm, which is equivalent to the effluent flow from each septum.

An operating problem encountered was the decrease in effluent quality due to the pulsing of spent carbon from an off-line adsorber. The two adsorbers remaining on-line experience a

"shock" as the individual established flow rates increased to include that portion of flow previously handled by the third adsorber. The decrease in effluent quality was due to the expelling of oil and carbon fines from the adsorber when flow was restored. In order to reduce the "shock" to the carbon bed, the times required to open and close were set to maintain the established flow rate to each on-stream adsorber during the carbon pulse cycle.

The time period for a "carbon out" cycle is a function of the time required to backwash the septums and the time required to return full flow to the adsorber while preventing the discharge of carbon particles with effluent. The maximum time an adsorber was off line during a "carbon out" cycle was two hours.

Excessive carbon losses have occurred when an adsorber is overpressured and vents to the atmosphere. In order to retain This carbon, the adsorbers are now operated as pressure vessels. Rupture discs and vacuum relief valves were installed in the adsorber vent line. A pressure sensor was installed in the vent line and signals for an automatic shutdown in the event an adsorber is overpressured.

After one year of operation, thickness measurements were made on the adsorber vessels using a sonoray instrument. Concentrated measurements were made on the lower cone of each adsorber. All measurements indicated no appreciable loss in wall thickness. Subsequent to two years of operation, leaks developed in the lower cone of each adsorber. External repairs were made by securing a rubber gasket and a piece of rolled steel over the leaking area.

SECTION 9

ECONOMIC EVALUATION

Table 15 presents the Wastewater Treatment Plant capital cost which totaled $1,812,000. This figure is reported in 1973 dollars.

Tables 16-19 present detailed operating costs for the filters, adsorbers, carbon regenerator, and solids dewatering system. The total annual operating cost for the period October, 1973-September, 1974, was $233,980 (Table 20). The carbon regeneration system accounted for 46 percent of this operating cost. The unit cost of regeneration including carbon makeup was 8.87¢/# carbon regenerated. The unit cost without including carbon makeup was 5.84¢/# carbon regenerated. The filters accounted for 10 percent of the annual operating cost, the adsorbers 17 percent and the solids dewatering system accounted for 27 percent.

Table 21 summarizes project capital and operating costs. In all systems, operating labor and utilities accounted for the majority of the operating costs. Operating labor accounted for 44 percent of the annual operating costs while utilities accounted for 25 percent. Makeup carbon for the regeneration system accounted for 16 percent of the annual operating costs. Maintenance accounted for approximately 10 percent.

Major needs contributing to maintenance costs were resurfacing of the scroll centrifuge internal flights, modifications to the solids dewatering feed system to eliminate clogging problems, addition of structural support to the carbon adsorber skirts and replacement of the carbon adsorber effluent septums to correct the septum bending problem.

TABLE 15

Wastewater Treatment Plant
Capital Cost

Item	Cost
Engineering	$ 125,000
Surge Basin	33,900
Sand Filters	187,100
Carbon System	348,200
Carbon Charge	111,200
Solids Dewatering	152,300
Tanks	93,800
Pumps	13,400
Building	149,200
Piping	179,600
Electrical	199,900
Instrumentation	137,500
Structural	22,100
Foundations	45,900
Concrete	12,800
Total	$1,812,000

TABLE 16

Dual Media Filters
Operating Costs
10/1/73 - 9/30/74

	Basis	Cost
Operating Labor	Refinery Records (12% of total WTP)	$12,340
Power	79 HP ($78.68 HP/YR)	6,220
Instrument Air	21 SCFM @0.03/1000 SCF (35% of total WTP)	330
Steam	Steam Tracing	350
Maintenance	Refinery Records	560
Supplies	Refinery Records	400
		$20,200

TABLE 17

Activated Carbon Adsorption
Operating Costs
10/1/73 - 9/30/74

	Basis	Cost
Operating Labor	Refinery Records (25% of total WTP)	$24,690
Power	115 HP ($78,68/HP/YR)	9,020
Instrument Air	6 SCFM @ $0.03/1000 SCF (10% of total WTP)	100
Steam	Steam Tracing	400
Maintenance	Refinery Records	3,180
Supplies	Refinery Records	400
		$37,790

TABLE 18

Activated Carbon Regeneration
Operating Costs
1,195,920# Regenerated
10/1/73 - 9/30/74

	Basis	Cost
Operating Labor	Refinery Records (38% of total WTP)	$37,030
Carbon Makeup	86,020# 42¢/lb. carbon cost (equates to 7.2% makeup)	36,130
Power	31.5 HP ($78.68/HP/YR)	2,470
Instrument Air	21 SCFM @ $0.03/1000 SCF (35% of total WTP)	330
Steam	1# steam/#carbon regen. $2.75/1000# steam + $560 Steam Tracing	3,850
Fuel	Refinery Records Based on Regen. Rate $0.80/MM BTU 20,740 MM BTU	16,590
Maintenance	Refinery Records	9,260
Supplies	Refinery Records	400
		$106,060
	Unit Costs Including Makeup	8.87¢/# carbon
	Unit Costs Not Including Makeup	5.84¢/# carbon

TABLE 19

Solids Dewatering
Operating Costs
10/1/73 - 9/30/74

	Basis	Cost
Operating Labor	Refinery Records (25% of total WTP)	$24,690
Power	149 HP ($78.68/HP/YR)	11,700
Instrument Air	12 SCFM @ $0.03/1000 SCF (20% of total WTP)	200
Steam	Steam Tracing and Heat Exchangers	3,850
Maintenance	Refinery Records	8,580
Supplies	Refinery Records	400
Sludge Disposal	Refinery Records	10,510
		$59,930

TABLE 20

Wastewater Treatment Plant

Annual Operating Costs

	Annual Operating Cost	% Of Total
Dual Media Filtration	$ 20,200	10
Activated Carbon Adsorption	37,790	17
Carbon Regeneration	106,060	46
Solids Dewatering	59,930	27
TOTAL	223,980	100

TABLE 21

Wastewater Treatment Plant

Project Cost Summary

Item	Cost	% of Total
Capital	1,812,000	-
Operating		
Labor	98,750	44
Utilities	55,410	25
Carbon Makeup	36,130	16
Sludge Disposal	10,510	5
Maintenance	21,580	10
Supplies	1,600	<1
TOTAL	223,980	100

APPENDIX A

WASTEWATER CONCENTRATION HISTOGRAMS

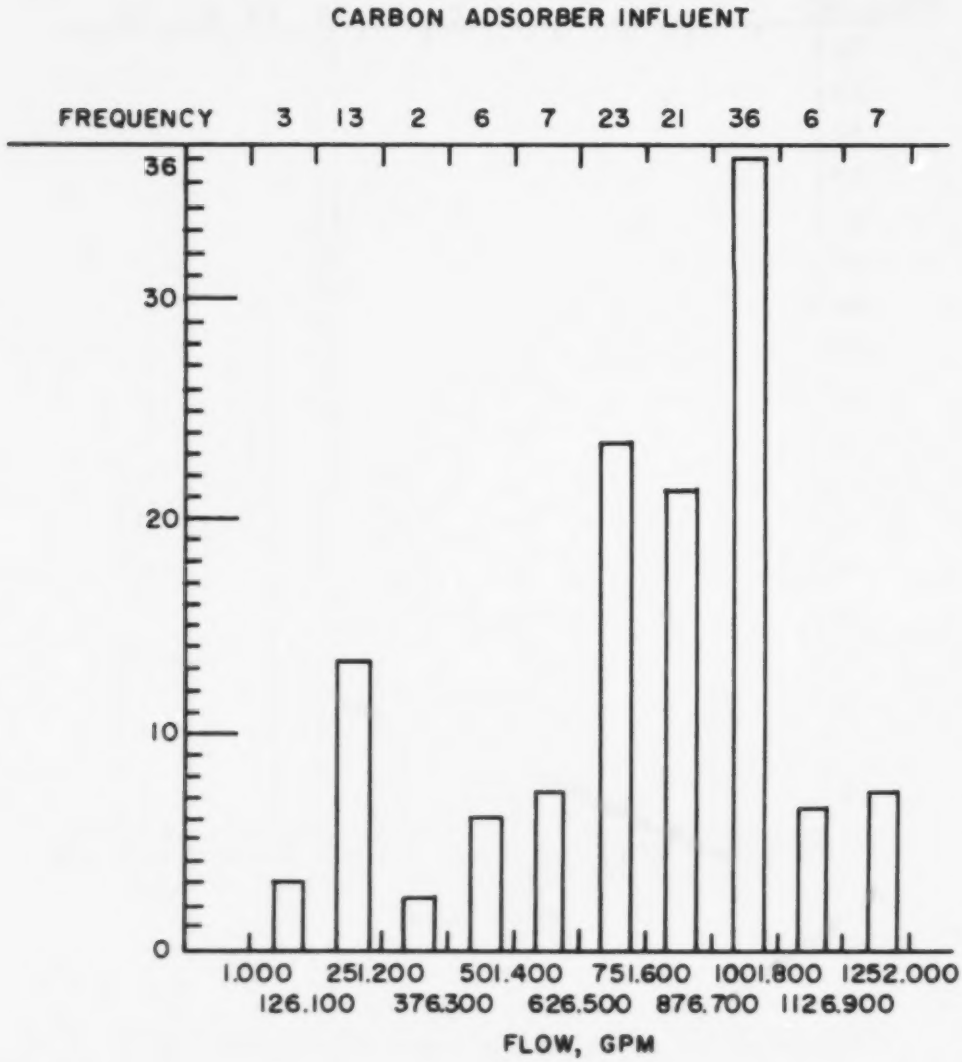


Figure A-1. Flow Histogram

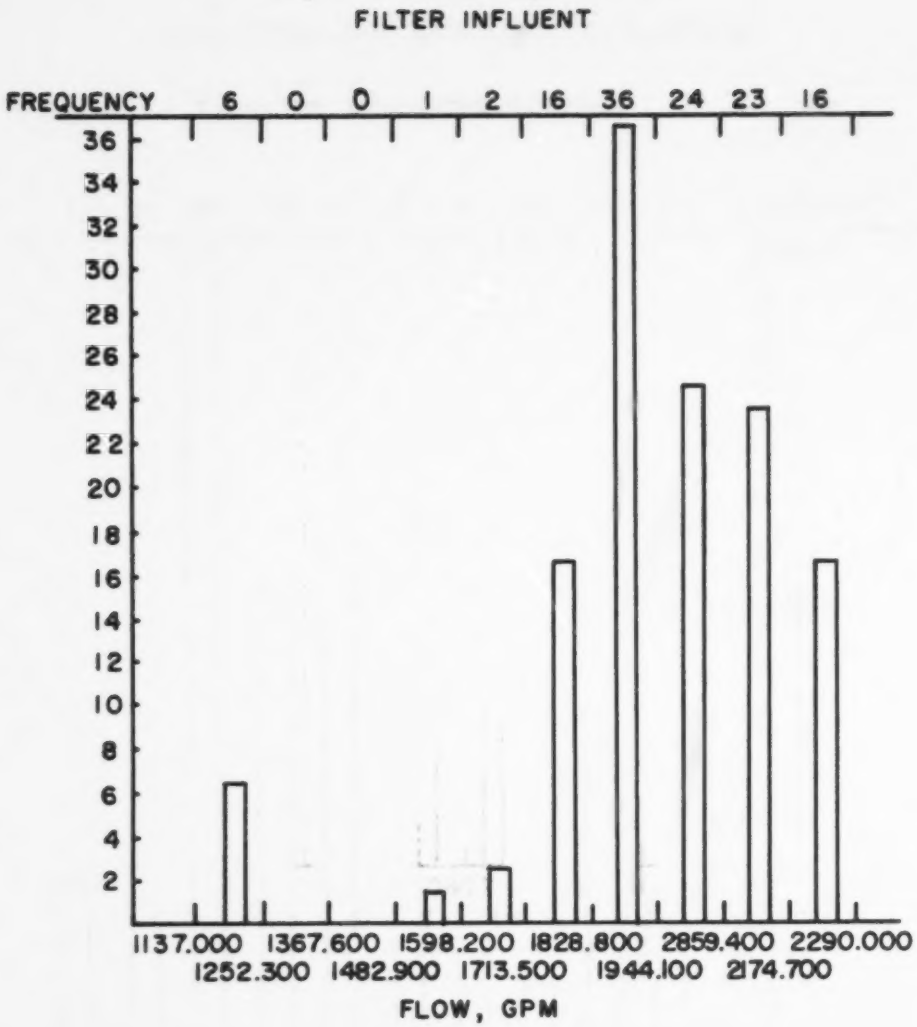


Figure A-2. Flow Histogram, filter influent

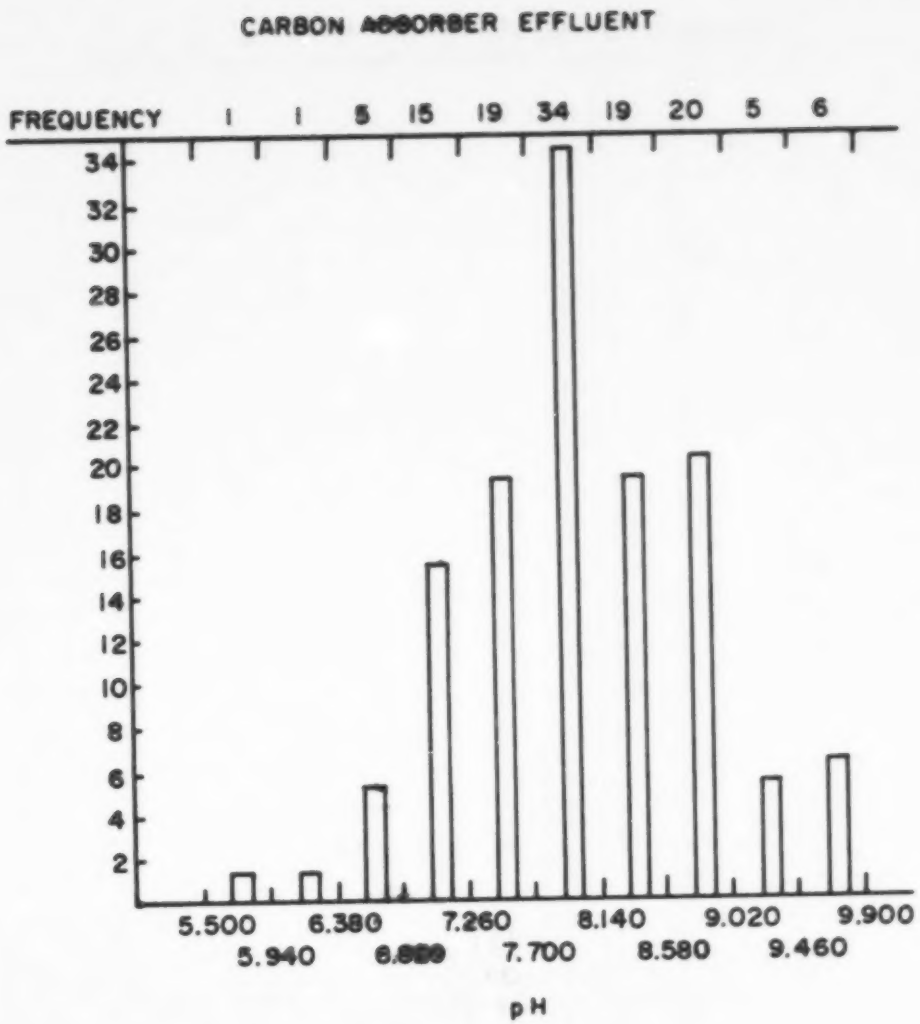


Figure A-3. pH Histogram, carbon adsorber effluent

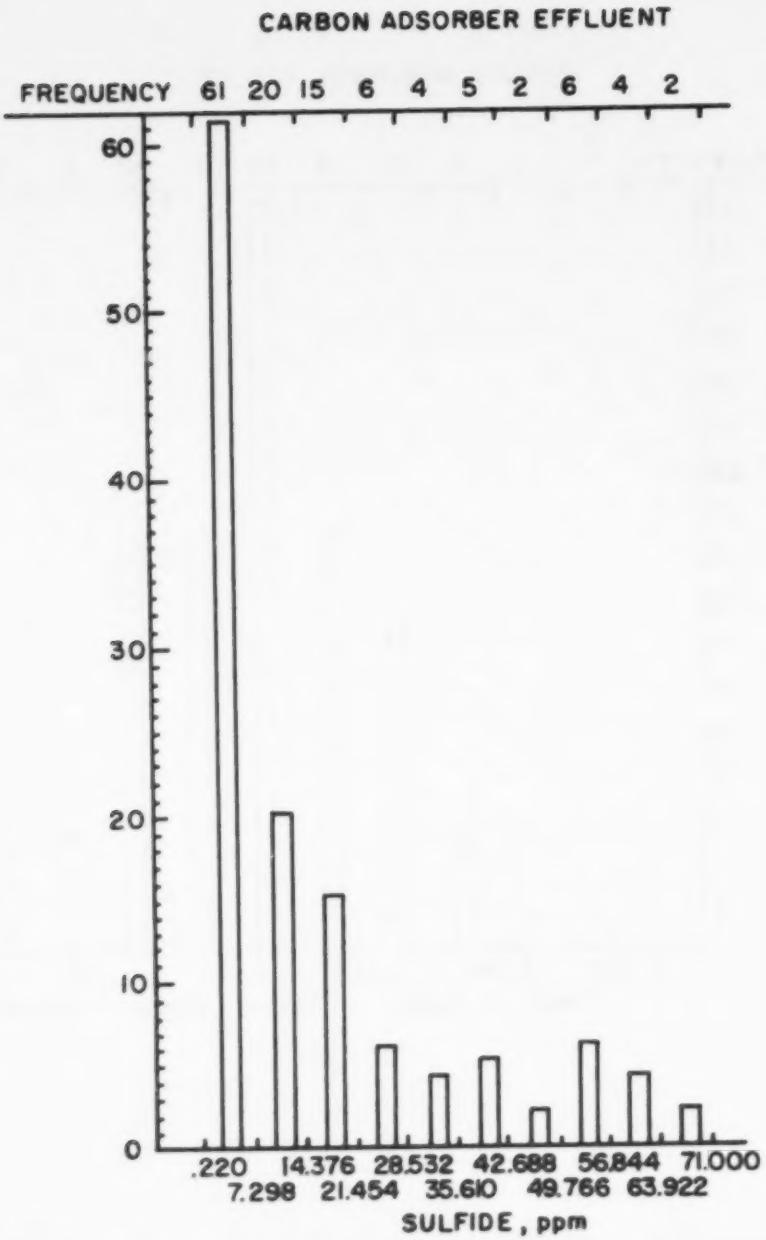


Figure A-4. Sulfide Histogram

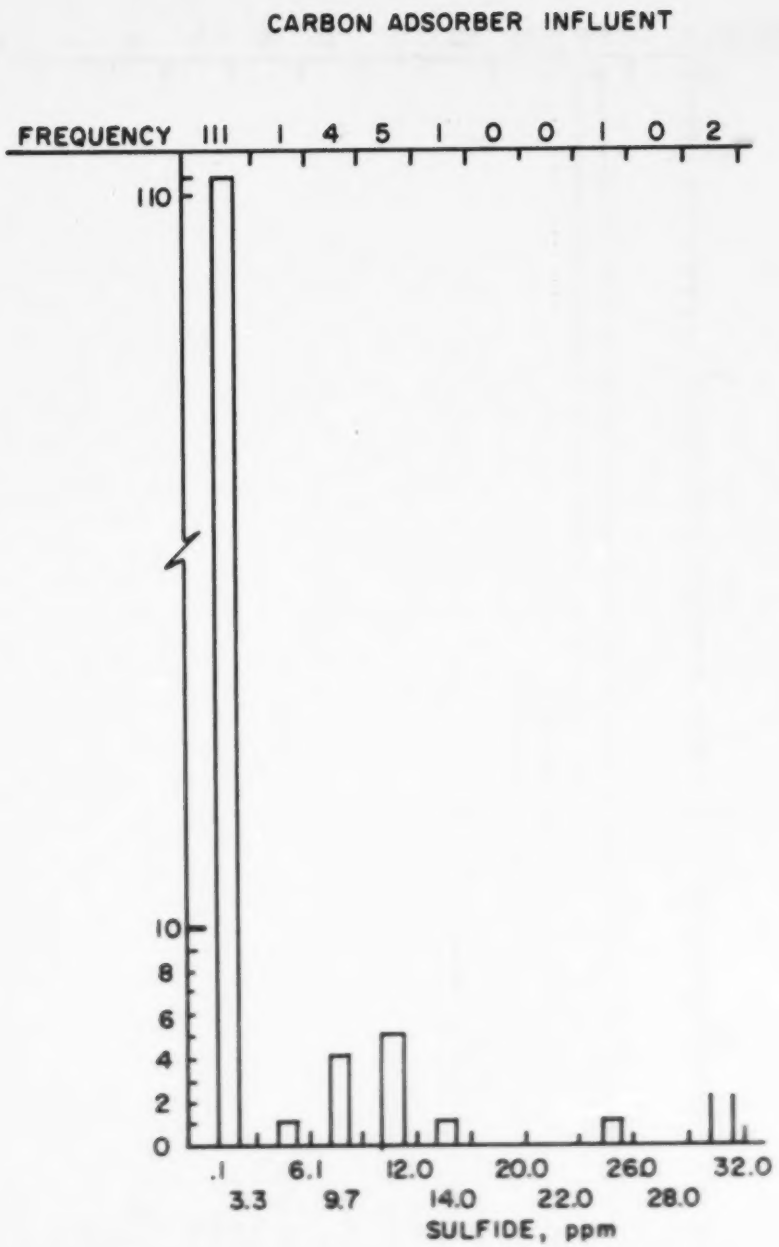


Figure A-5. Sulfide Histogram

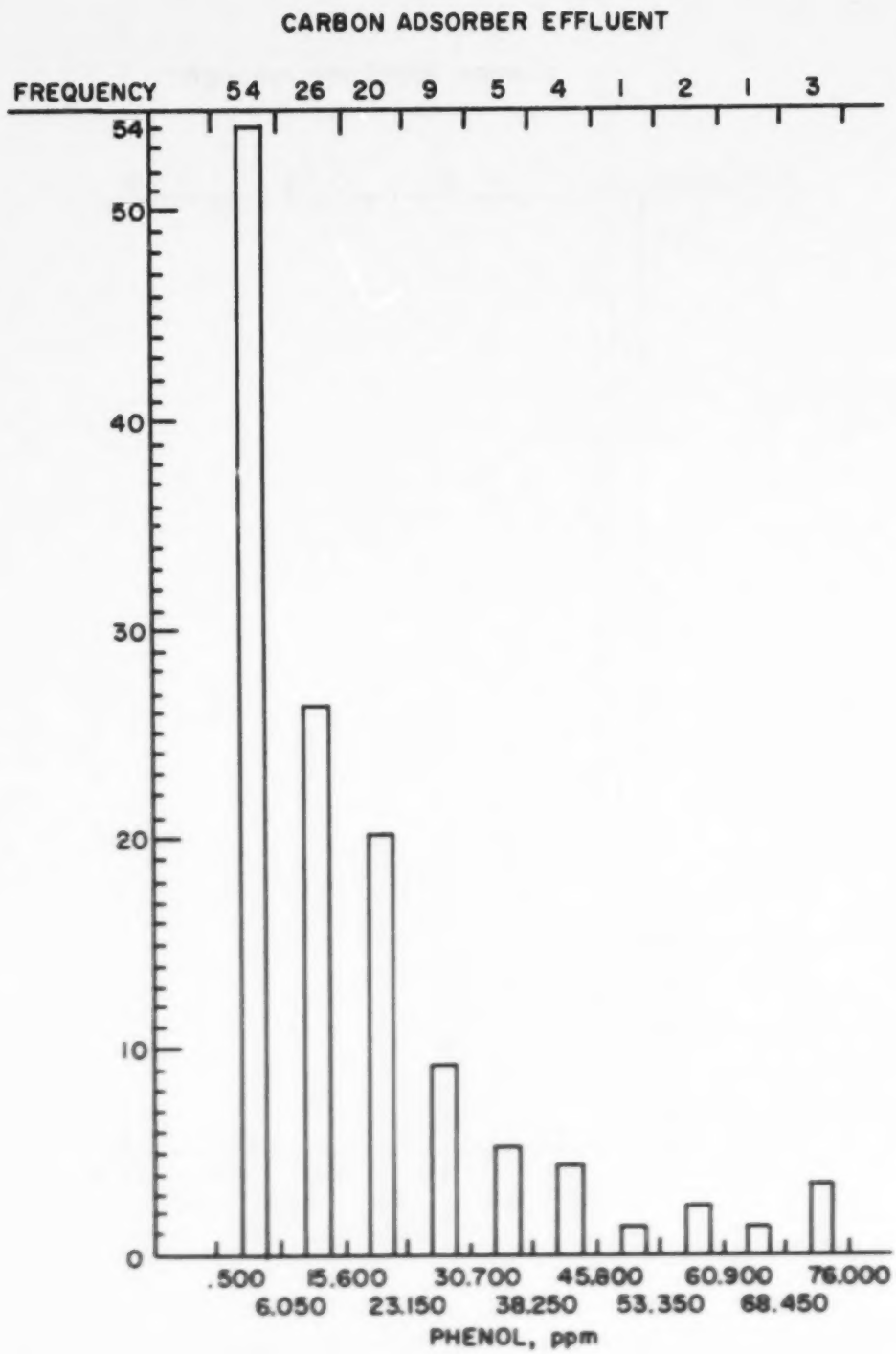


Figure A-6. Phenol Histogram

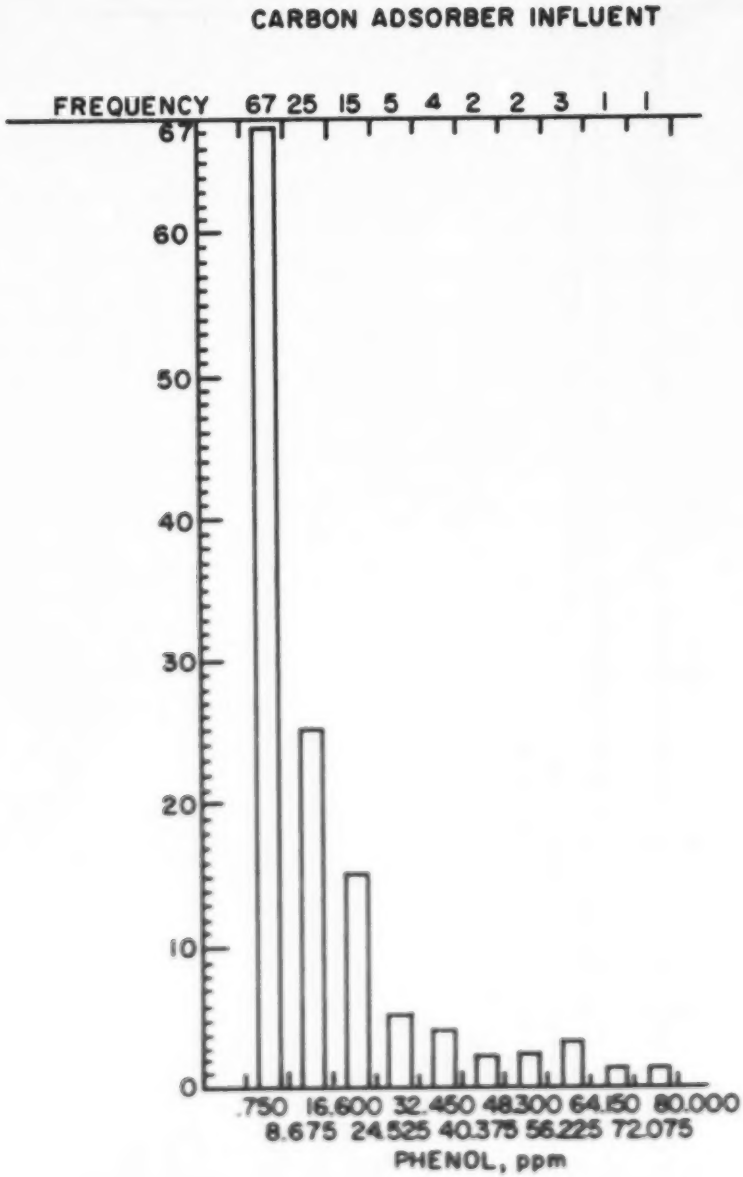


Figure A-7. Phenol Histogram, filter effluent

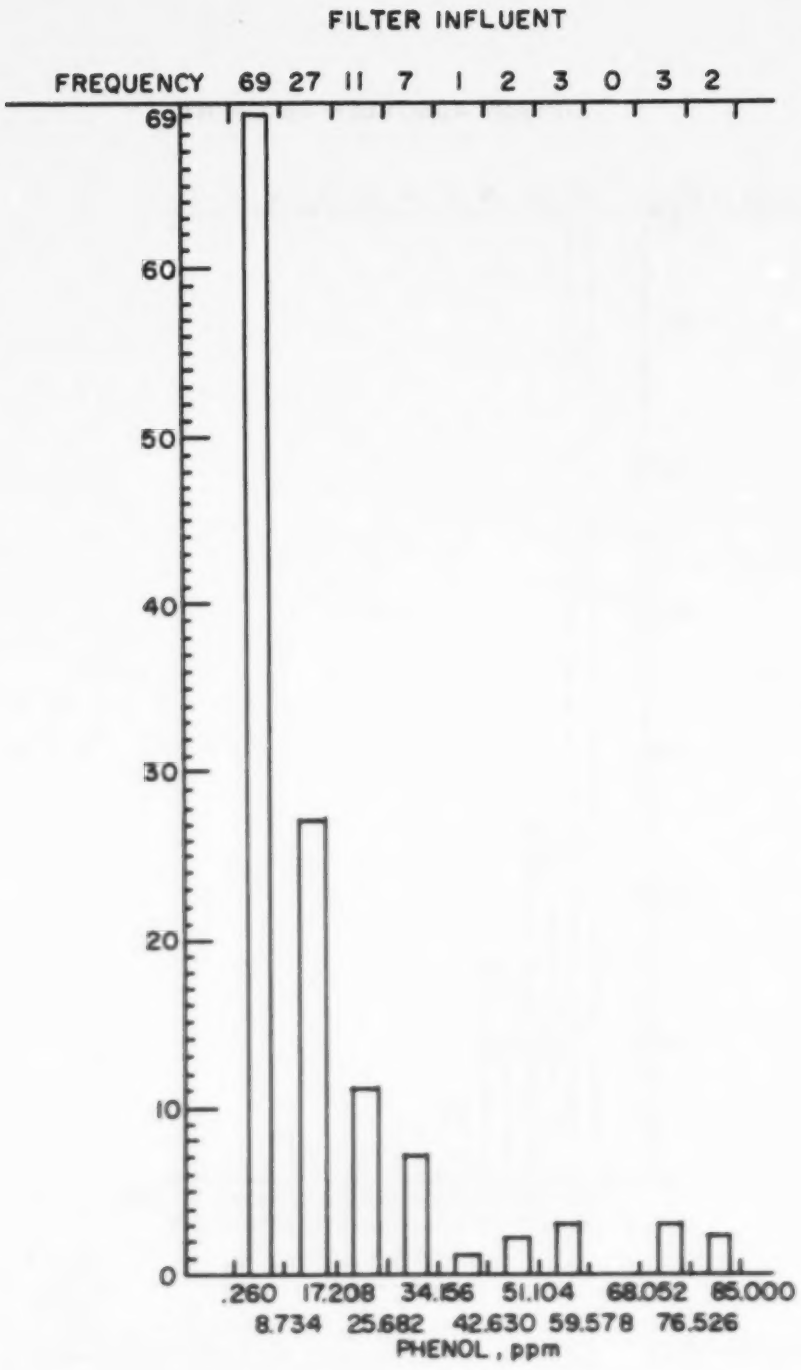


Figure A-8. Phenol Histogram

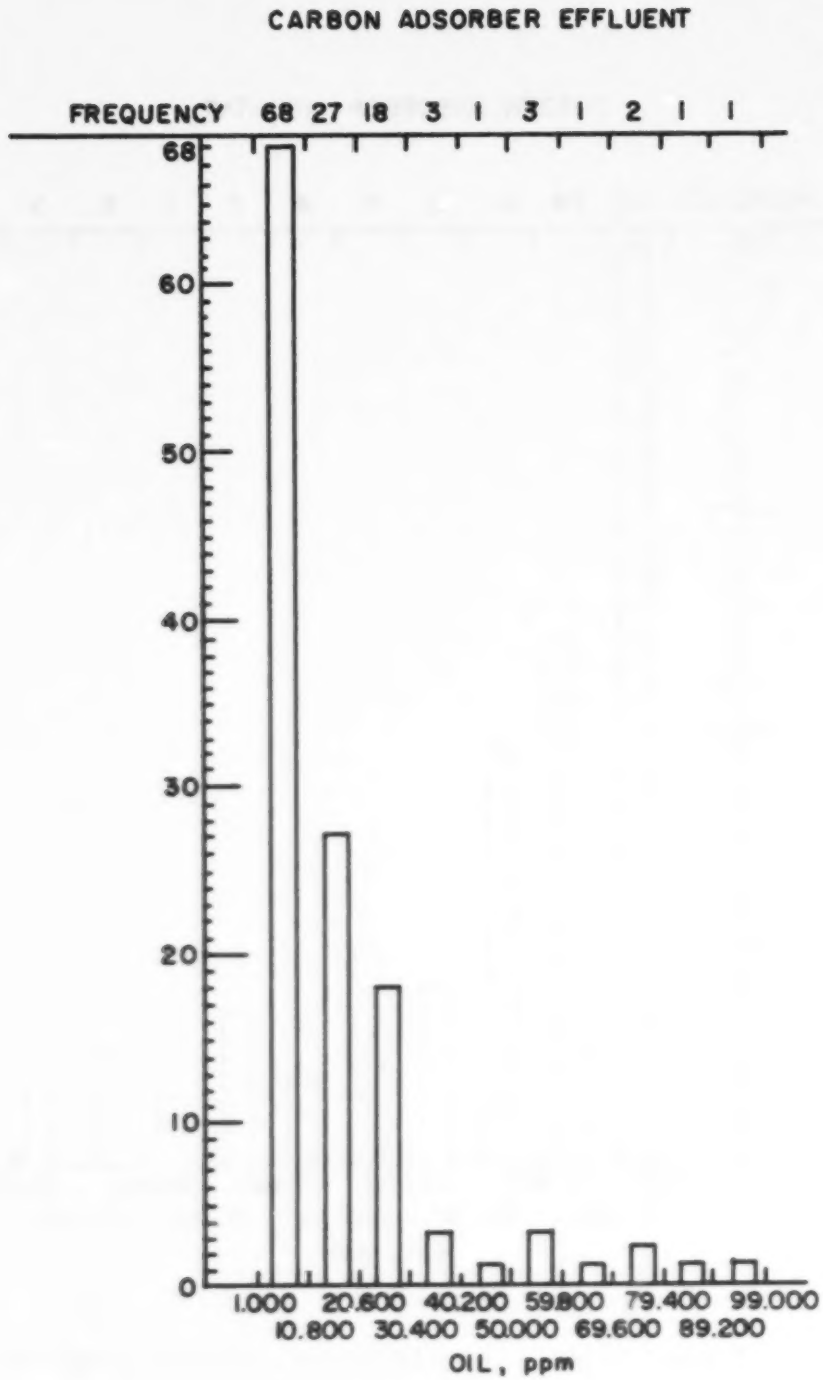


Figure A-9. Oil Histogram

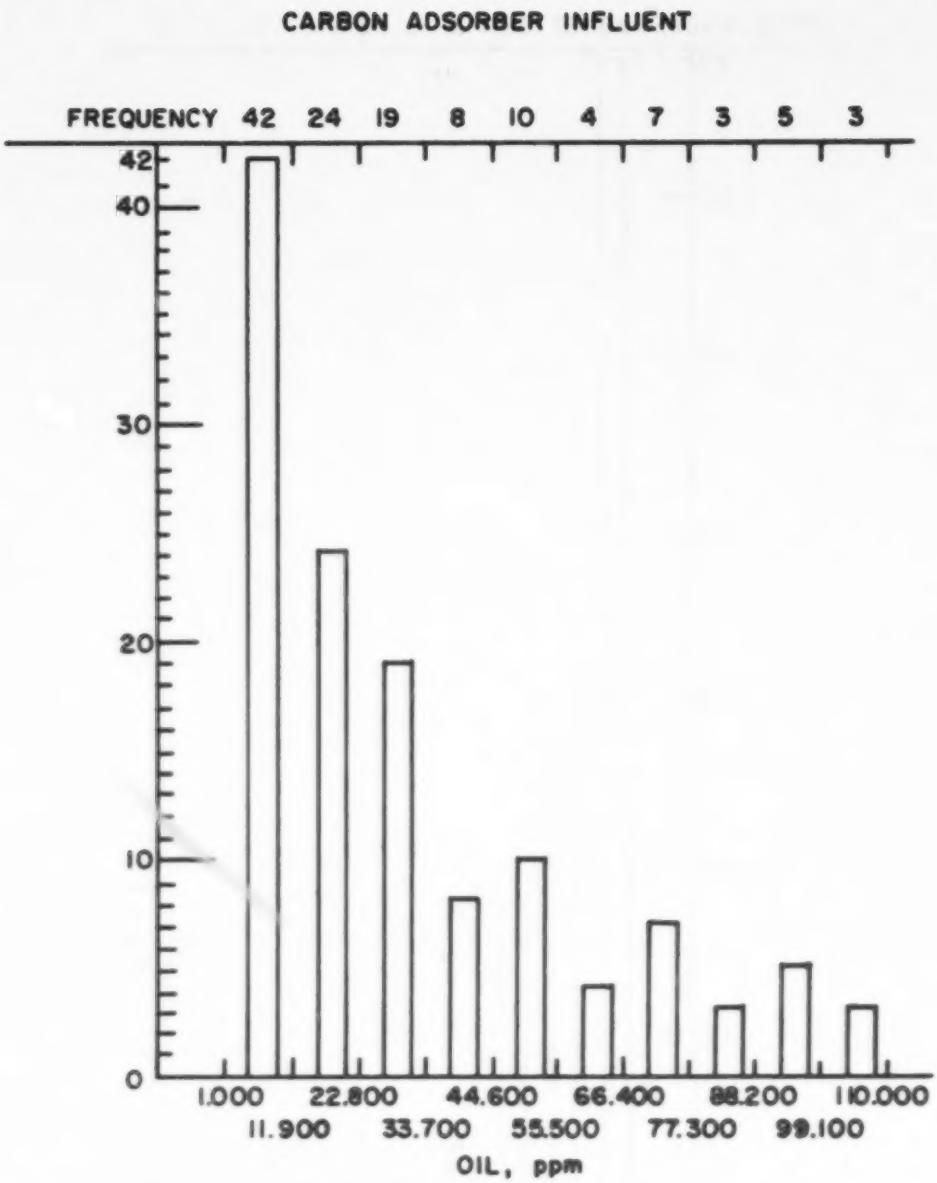


Figure A-10. Oil Histogram, filter effluent

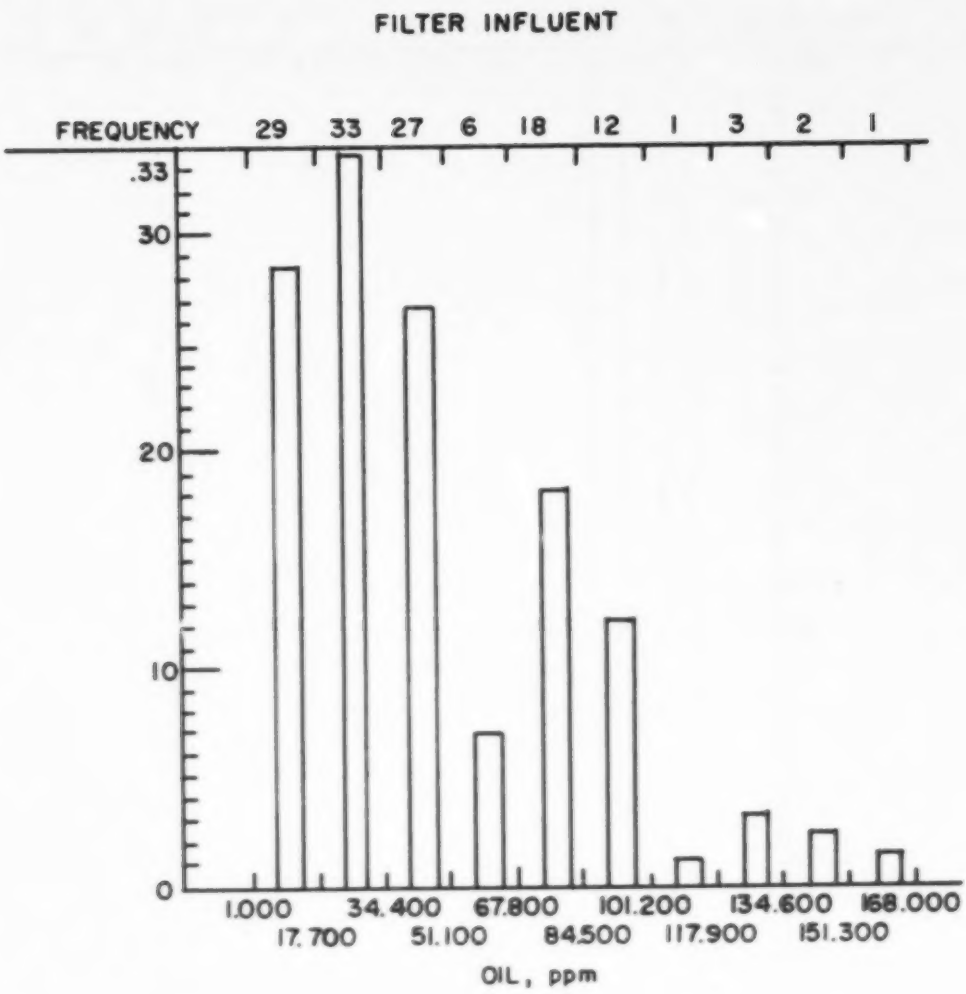


Figure A-11. Oil Histogram

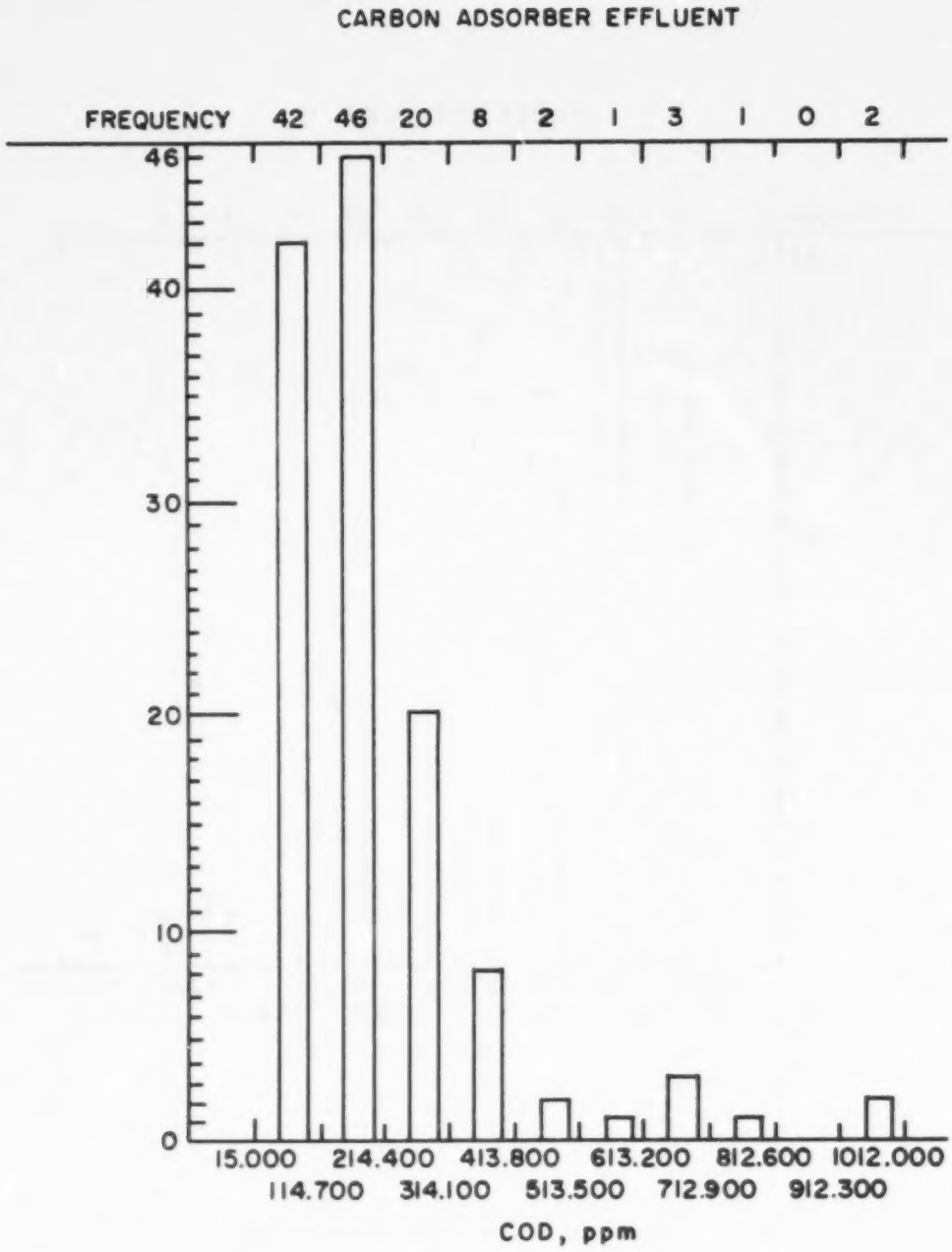


Figure A-12. COD Histogram

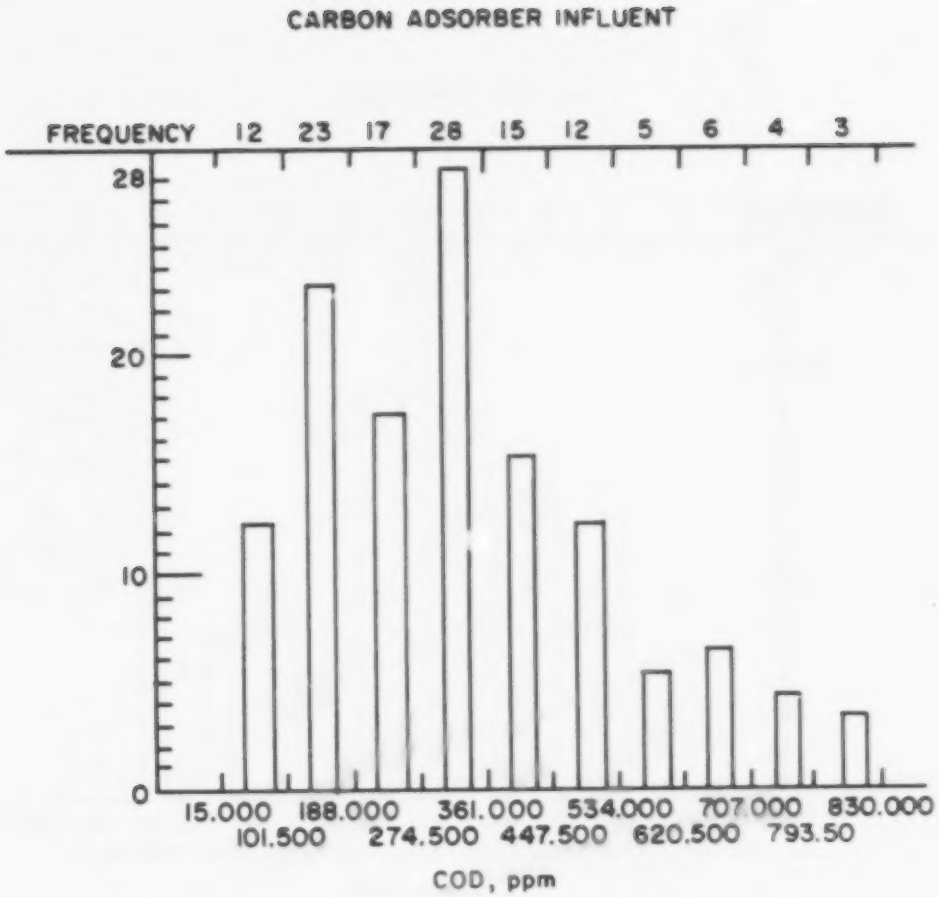


Figure A-13. COD Histogram, filter effluent

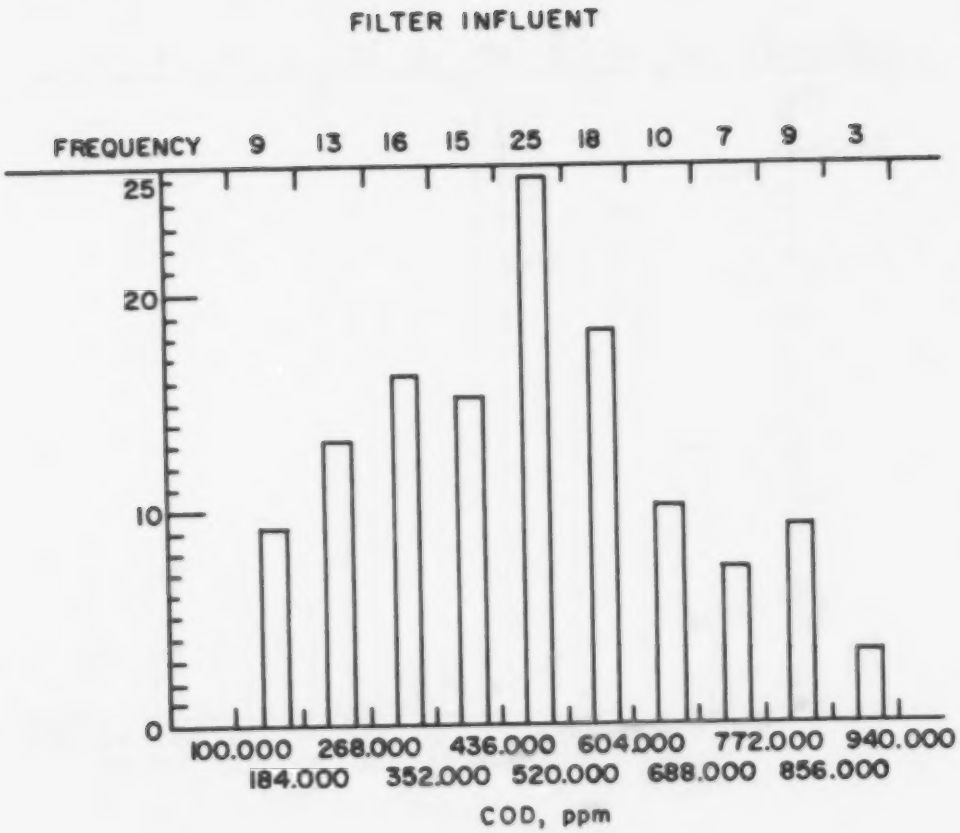


Figure A-14. COD Histogram

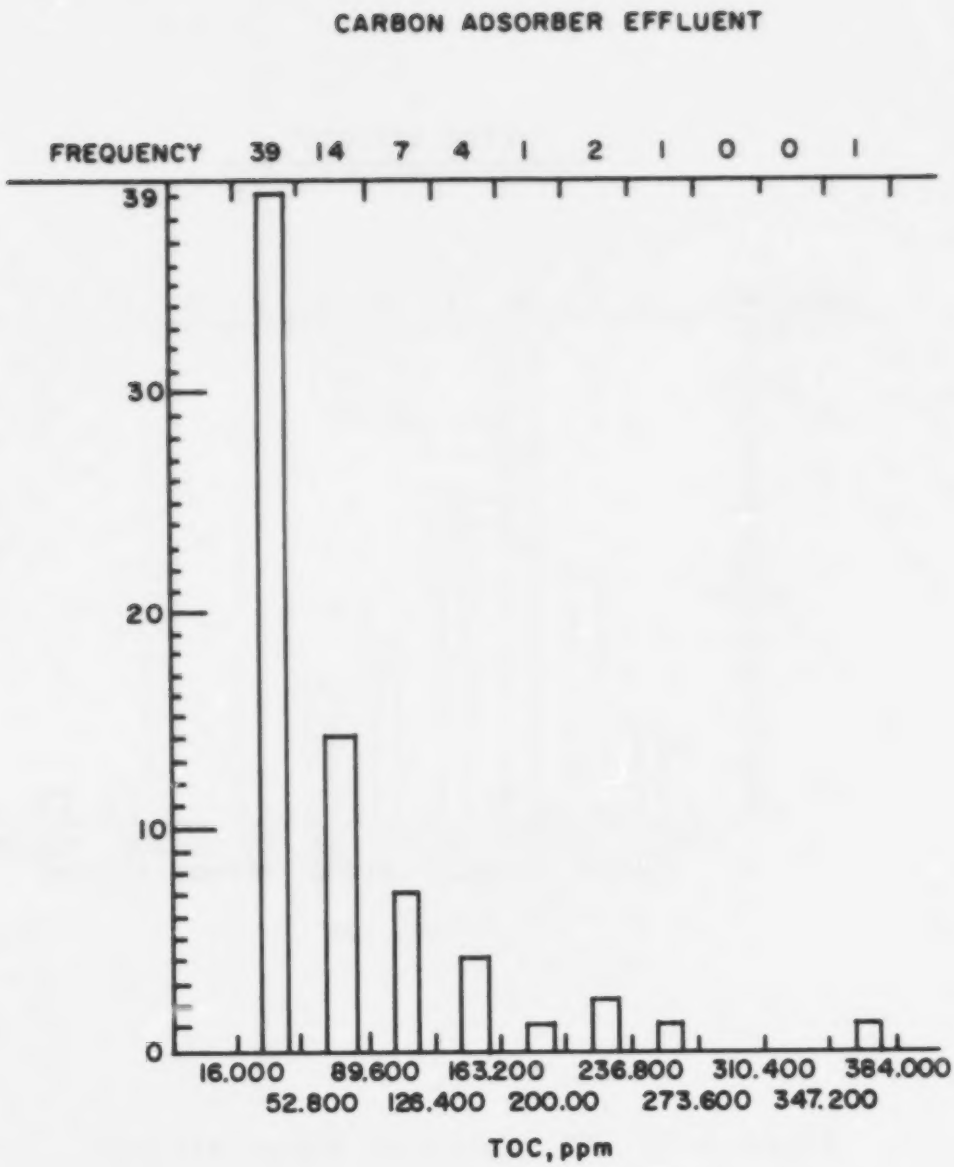


Figure A-15. TOC Histogram

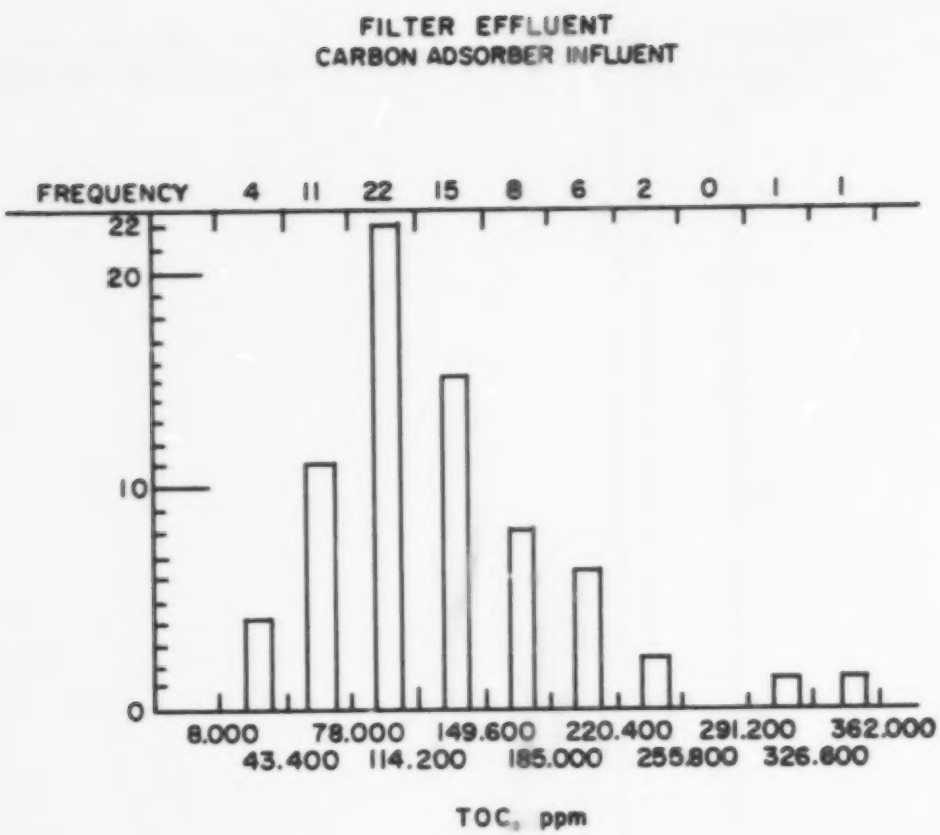


Figure A-16. TOC Histogram, filter effluent

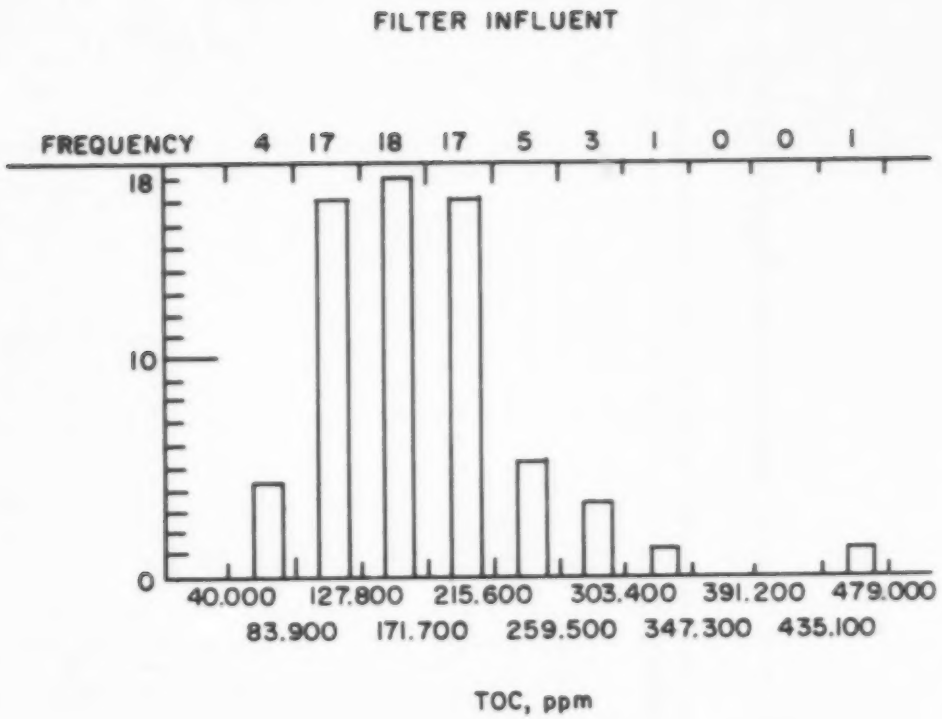


Figure A-17. TOC Histogram

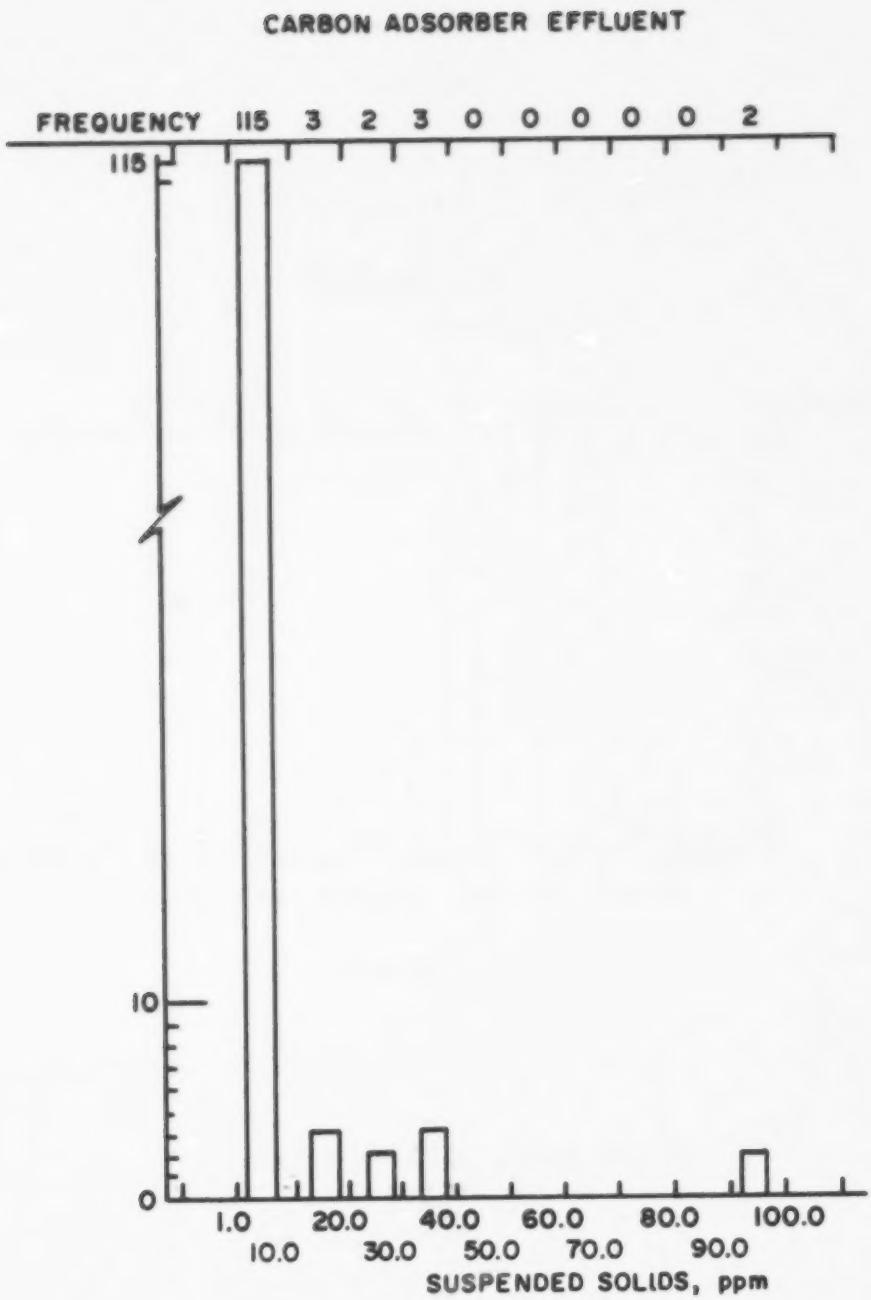


Figure A-18. Suspended Solids Histogram

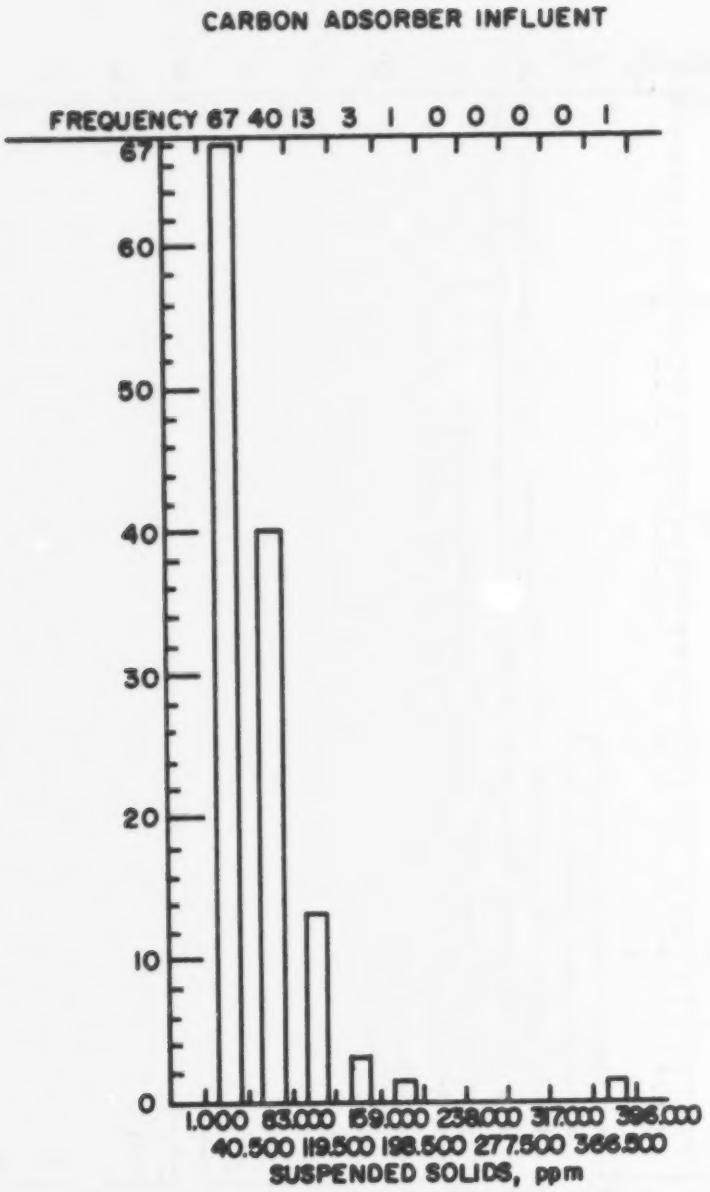


Figure A-19. Suspended Solids Histogram, filter effluent

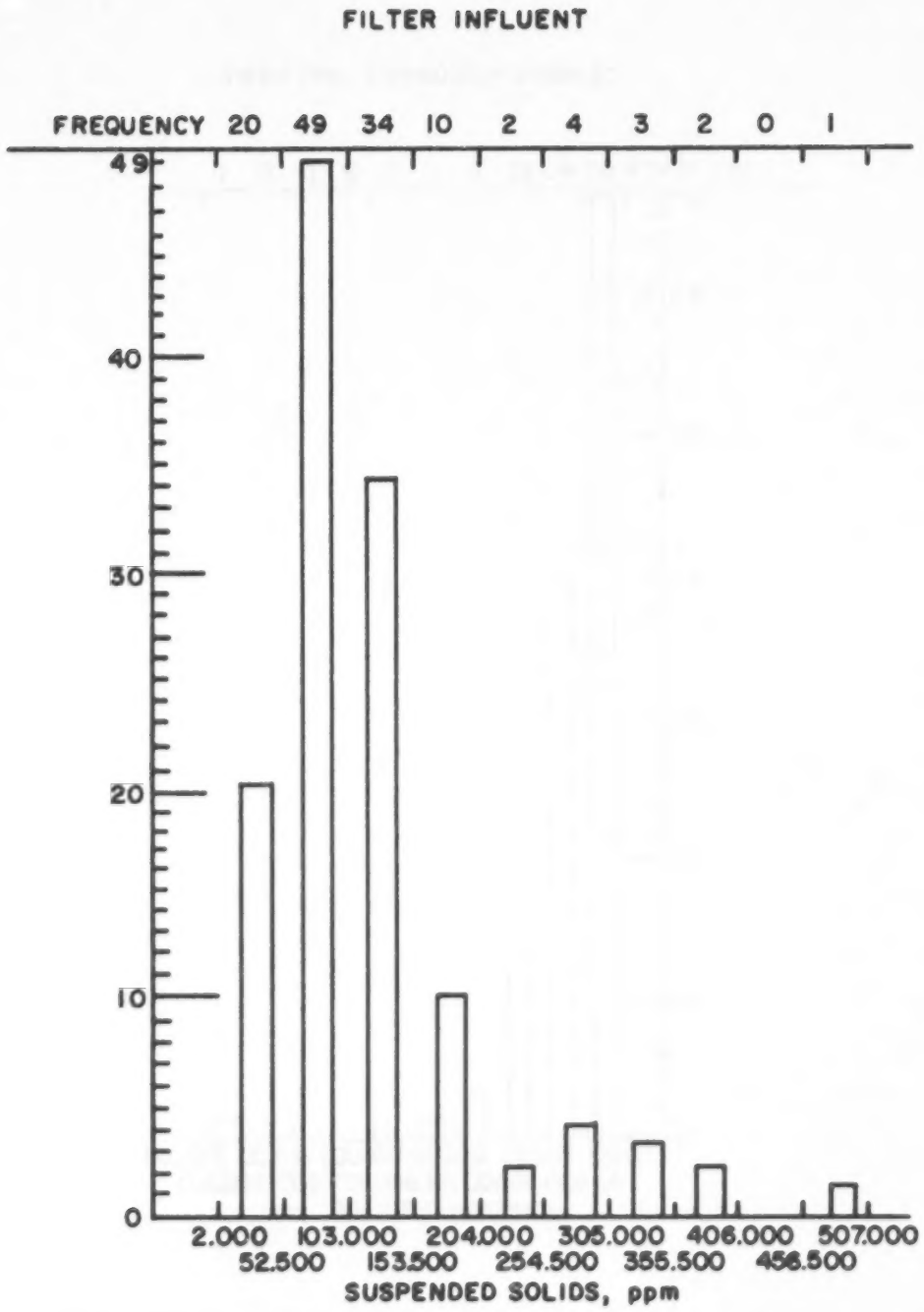


Figure A-20. Suspended Solids Histogram

APPENDIX B

ENGLISH-TO-METRIC UNIT CONVERSIONS

Multiply This	By This	To Obtain This	
lbs	0.4536	kg	kilograms
short tons	0.9072	metric tons	metric tons (1000 kg)
short tons	907.2	kg	kilograms
inches	2.54	cm	centimetres
feet	0.3048	m	metres
statute miles	1.609	km	kilometres
gallons	3.785	l	litres (1000 litres = 1 m^3)
barrels	0.1590	m^3	cubic metres
Btu	0.252	kcal	kilocalories
SCF	0.02679	nm^3	normal cubic metres
Btu/lb	0.5556	kcal/kg	kilocalories/kilogram
Btu/CF	8.899	kcal/m^3	kilocalories/cubic metre
Btu/SCF	9.406	kcal/nm^3	kilocalories/normal cubic metre
10^9 Btu/day	252	Gcal/day	gigacalories/day
10^6 Btu/day	252	Mcal/day	megacalories/day
MM Btu/hr	252	Mcal/hr	megacalories/hour
SCFD	0.02679	nm^3/day	normal cubic metres/day
MM SCFD	0.02679	10^6 nm^3/day (Mnm^3/day)	million normal cubic metres/day (mega normal cubic metres/day)
SCF/MM Btu	0.1063	nm^3/Gcal	normal cubic metres/gigacalorie
lbs/MM Btu	1.8	kg/Gcal	kilograms/gigacalorie
lbs/CF	16.02	kg/m^3	kilograms/cubic metre
psi	0.07031	kg/cm^2	kilograms/square centimetre
gpm	0.227	m^3/hr	cubic metres/hour
acre-ft/year	0.1408	m^3/hr	cubic metres/hour
horsepower	745.7	W	watts
nautical miles	1.852	km	kilometres
knot	1.852	km/hr	kilometres/hour

(a) A SCF of gas is measured at 60°F and atmospheric pressure, and a nm^3 of gas is measured at 0°C and atmospheric pressure.

(b)

Exponential	English	SI Metric
10^3	M or thousand	k or kilo
10^6	MM or million	M or mega
10^9	billion (U.S.)	G or giga
10^{12}	billion (U.K.)	T or tera